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EVALUATION OF THE FLUIDIZED-BED COMBUSTION PROCESS
VOLUME I. PRESSURIZED-BED COMBUSTION PROCESS DEVELOPMENT AND EVALUATION

WESTINGHOUSE RESEARCH LABORATORIES

PREPARED FOR
ENVIRONMENTAL PROTECTION AGENCY

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EVALUATION OF THE FLUIDIZED-BED COMBUSTION PROCESS

VOLUME I -

PRESSURIZED FLUIDIZED-BED COMBUSTION PROCESS DEVELOPMENT AND EVALUATION

by

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In particular, we want here to express our high regard for and acknowledge the contribution of personnel at Westinghouse Research Laboratories and the personnel at ORD who conceived the overall fluidized bed combustion boiler effort and who have defined, monitored, and supported the efforts of Westinghouse and others on the program. Mr. P. P. Turner, Chief of the Advanced Process Section, has served as EPA project officer on our work. Numerous enlightening and helpful discussions have been held with Mr. Turner; with section members D. Bruce Henschel and Sam Rakes; and with R. P. Hangebrauck, Chief of the Demonstration Projects Branch. Personnel from the Westinghouse Research Laboratories have made significant contributions. Mr. E. J. Vidt participated in the evaluation. Mr. R. E. Brinza and Mr. C. Spangler assisted in the collection of data. Drs. L. E. Brecher and C. R. Wolfe participated in the design and development of the pressurized thermogravimetric analysis system. Mr. W. F. Kittle contributed to each stage of the design, development and use of the pressurized thermogravimetric analysis system. Westinghouse Power Generation Systems, Gas Turbine Systems Division, Heat Transfer Division, Power Generation Services, and Instrumentation Systems Division personnel were consulted and participated in the evaluation.

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PREFACE

The Office of Research and Development (ORD) of the United States Environmental Protection Agency has organized and is sponsoring a fluidized bed fuel processing program. Its purpose is to develop and demonstrate new methods for utilizing fossil fuels -- particularly coal and oil -- in utility power plants. These methods should:

- Meet environmental goals for SO_2, NO_x, ash, smoke emissions, and wastes
- Compete economically with alternative means for meeting these abatement goals.

Westinghouse Research, under contract to the Office of Research and Development (ORD) of the Environmental Protection Agency, is carrying out a study to evaluate and develop fluidized bed combustion and oil gasification in air pollution abatement. The goals of this work are

- To identify which fluidized processes might be economically employed in utility power plants or industrial boilers to reduce SO_2, particulate, and NO_x emissions
- To assist in planning and implementing a program to develop the fluidized bed systems deemed effective in air pollution control and economical in steam/power production.

Tasks in the evaluation of fluidized bed combustion set forth y EPA which have been completed at Westinghouse under a previous contract are

- To search the technical and patent literature in fluidized bed combustion, to canvass commercial organizations with expertise pertaining to this field, and to survey the market for industrial boilers and utility power systems

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- To design a fluidized bed industrial boiler and two fluidized bed utility power systems -- one 300 MW capacity, the other 600 MW; and to provide performance and cost projections for the equipment designed
- To provide technical consultation and assistance on the ORD fluidized bed fuel processing program, including both combustion and gasification
- To conceptualize a development plant which will prove the configurational or operating features of the fluidized bed boiler designs not confirmed by prior art or experience
- To assess the effectiveness and economics of an atmospheric-pressure, fluidized bed oil gasification-combustion system and to aid in planning for a demonstration installation of such a system.

The results of these surveys, designs, evaluations, and comparisons were published in a three-volume report, "Evaluation of the Fluidized Bed Combustion Process," in November 1971 under contract No. CPA 70-9.

These results provided the basis for the work performed from July 1971 to May 1973 which is described in this four-volume report. Tasks in the evaluation of the fluidized bed combustion process set forth by EPA under contract 68-02-0217 have focused on the development of pressurized fluidized bed combustion for power generation and fluidized bed oil gasification for power generation. These tasks, which are presented here, have included:

- Extensive process evaluation studies of the pressurized fluidized bed combustion system for power generation. These investigations predict the sensitivity of operating and design parameters selected for the base power plant design on plant economics; provide additional experimental data on sulfur removal and sorbent regeneration using limestones and dolomites; project economics and performance for various sulfur removal systems, both regeneration/sulfur recovery and once-through sorbent; establish a plant operation and control philosophy and evaluate alternative pressurized fluid bed combustion concepts, both economics and performance.

- Preparation of preliminary plans and a cost estimate for a 30 MW (equivalent) pressurized fluidized bed combustion boiler development plant. The design provides sufficient detail to locate a suitable plant site and to obtain a fixed cost bid for the preparation of detailed plans. The plant will provide the capability for studying the remaining technical problems effectively in order to achieve the greatest potential for reducing emissions and for generating economical electrical energy.
- Identification of a project team to demonstrate fluidized bed oil gasification/desulfurization for power generation. A cooperating utility -- New England Electric System -- has been identified to carry out a 50 MW demonstration plant program. Further process evaluation has been carried out and experimental data obtained on sulfur removal and spent stone disposal.
- Evaluation of pressurized oil gasification for combined cycle power generation. Oil gasification process concepts and options have been reviewed, material and energy balances projected, performance projected,and capital and energy costs estimated.
- Provision of technical consultation and assistance on the ORD fluidized bed fuel processing program, including both combustion and gasification processes. Technical and economic comparisons have been carried out on various fluidized bed fuel processing systems and various conventional means of steam/power generation.

This volume (Volume I) contains new data on pressurized fluid bed combustion and an evaluation of the current technology state. Volume II contains the appendices to this volume. Volume III presents the preliminary design of a 30 MW (equivalent) pressurized fluid bed boiler development plant. Volume IV includes the work on fluidized bed oil gasification/desulfurization.

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ALTERNATIVE PRESSURIZED FLUID BED BOILER CONCEPTS

P. Recirculating Bed Boiler Studies

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VOLUME III

PRESSURIZED FLUID BED BOILER DEVELOPMENT PLANT DESIGN

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VOLUME IV

FLUIDIZED BED OIL GASIFICATION/DESULFURIZATION

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- N. Partial Oxidation Processes
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CONVERSION FACTORS -- ENGLISH TO METRIC UNITS

	English System	Metric Equivalent
Length	in	2.54 cm
	ft	0.305 m
Area	in^2	6.45 cm^2
	ft^2	0.0930 m^2
Volume	in^3	16.39 cm^3
	ft^3	28.32 l
	gal	3.785 l
Mass	oz	28.35 gm
	lb	453.6 gm
	ton	907.2 kg
Pressure	lb/in^2	51.70 mm Hg
	in H_2O	1.865 mm Hg
Temperature	°F	1.8 (°C) + 32
	°R	1.8 °K
Energy	Btu	252 cal
Power	Btu/min	252 cal/min

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1. SUMMARY

Fluidized bed combustion technology can be applied with many different configurations and can be utilized in many different steam and power cycles.[1,2,3,4,5,6] Fluidized bed combustion systems, operating at elevated pressure in a combined cycle power plant, offer the greatest potential for producing electrical energy from fossil fuels within environmental constraints and at a cost less than conventional power plants using low-sulfur fuel or stack gas cleaning. This conclusion is based on a previous evaluation of the fluidized bed combustion process.[1] Further development and evaluation of pressurized fluid bed combustion systems have been carried out in order to assess more fully the concept potential. Results from the following tasks are reported:

- Economic sensitivity study. The effectiveness and economics of a pressurized fluid bed combustion boiler combined cycle power plant (Figure 1) using state-of-the-art power generation equipment (steam conditions 2400 psig/1000°F/1000°F, gas-turbine expander operating at 10:1 pressure ratio and 1600°F) were previously reported.[1] The economic sensitivity of the operating conditions and design parameters selected for this design was determined. The boiler design concept is presented in Figures 2 and 3, and the design parameters are summarized in Table 1.
- Sulfur removal system data. A pressurized thermogravimetric analysis system was used to collect kinetic data on sulfur dioxide sorption by limestones and dolomites and spent stone regeneration at projected operating conditions -- e.g., temperature, pressure, stone size, gas composition. These data provide information, not previously available, with which the desulfurization of fuels in a pressurized fluid bed system can be predicted and explained.

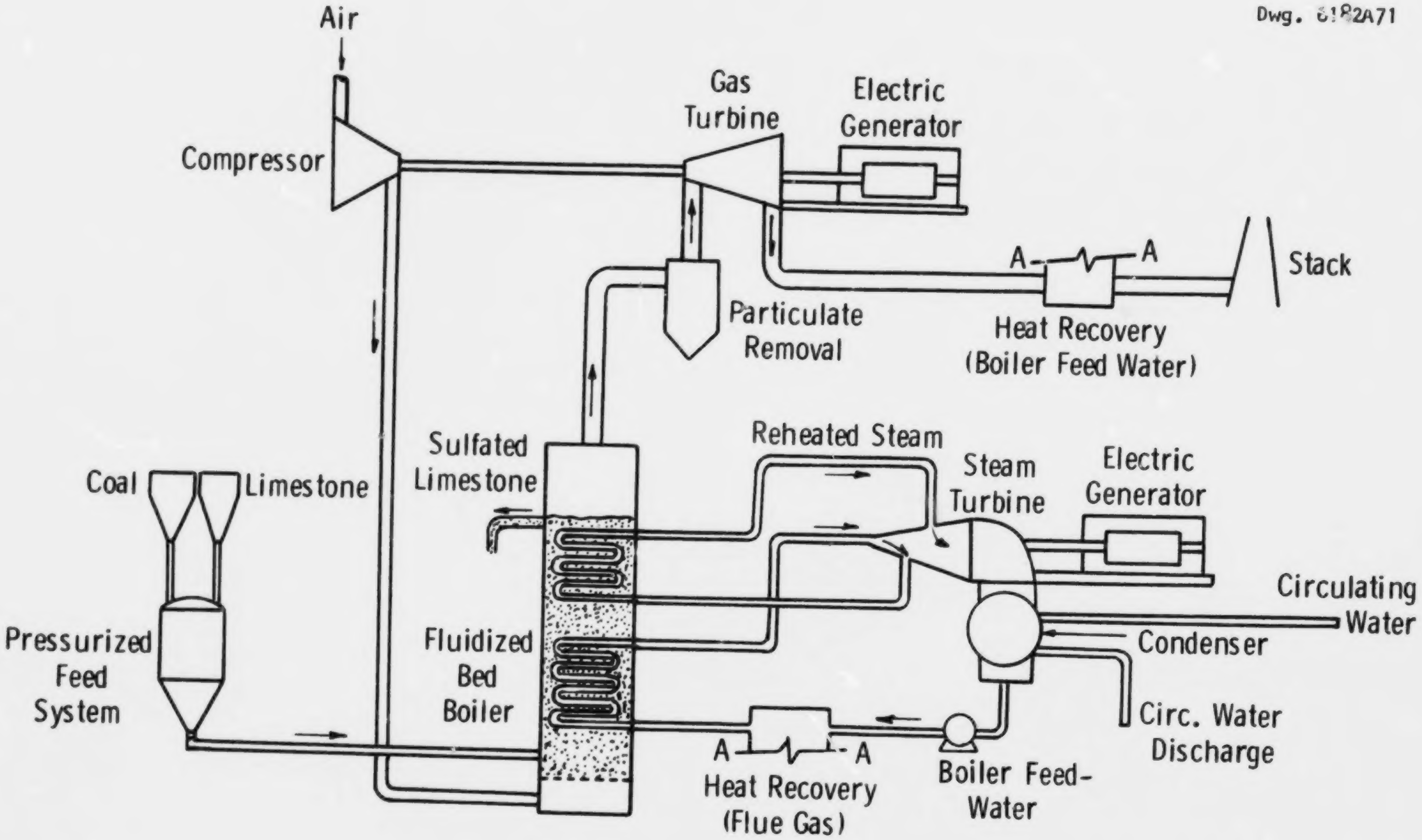
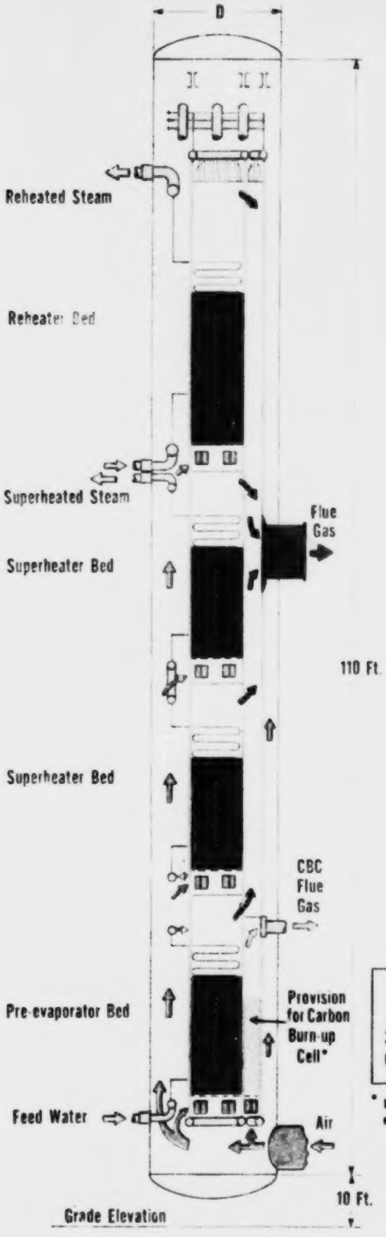

Dwg. 6182A71
Air
Compressor
Gas Turbine
Electric Generator
A
A
Stack
Particulate Removal
Heat Recovery (Boiler Feed Water)
Coal
Limestone
Sulfated Limestone
Reheated Steam
Steam Turbine
Electric Generator
Circulating Water
Condenser
Circ. Water Discharge
Pressurized Feed System
Fluidized Bed Boiler
A
A
Heat Recovery (Flue Gas)
Boiler Feed-Water

Figure 1–Pressurized Fluidized Bed Boiler Power Plant


D
Reheated Steam
Reheater Bed
Superheated Steam
Flue Gas
Superheater Bed
110 Ft.
Superheater Bed
CBC Flue Gas
Pre-evaporator Bed
Provision for Carbon Burn-up Cell*
Feed Water
Air
10 Ft.
Grade Elevation

PLANT SIZE	VESSEL DIAMETER, D
320 mw	12 Ft.
635 mw	17 Ft.

* Dependent on operating conditions excess air, bed depth, gas velocity

ELEVATION

PRESSURIZED FLUIDIZED BED STEAM GENERATOR
FOR COMBINED CYCLE PLANT
FOUR (4) REQUIRED

Figure 2 Pressurized Fluidized Bed Steam Generator Module

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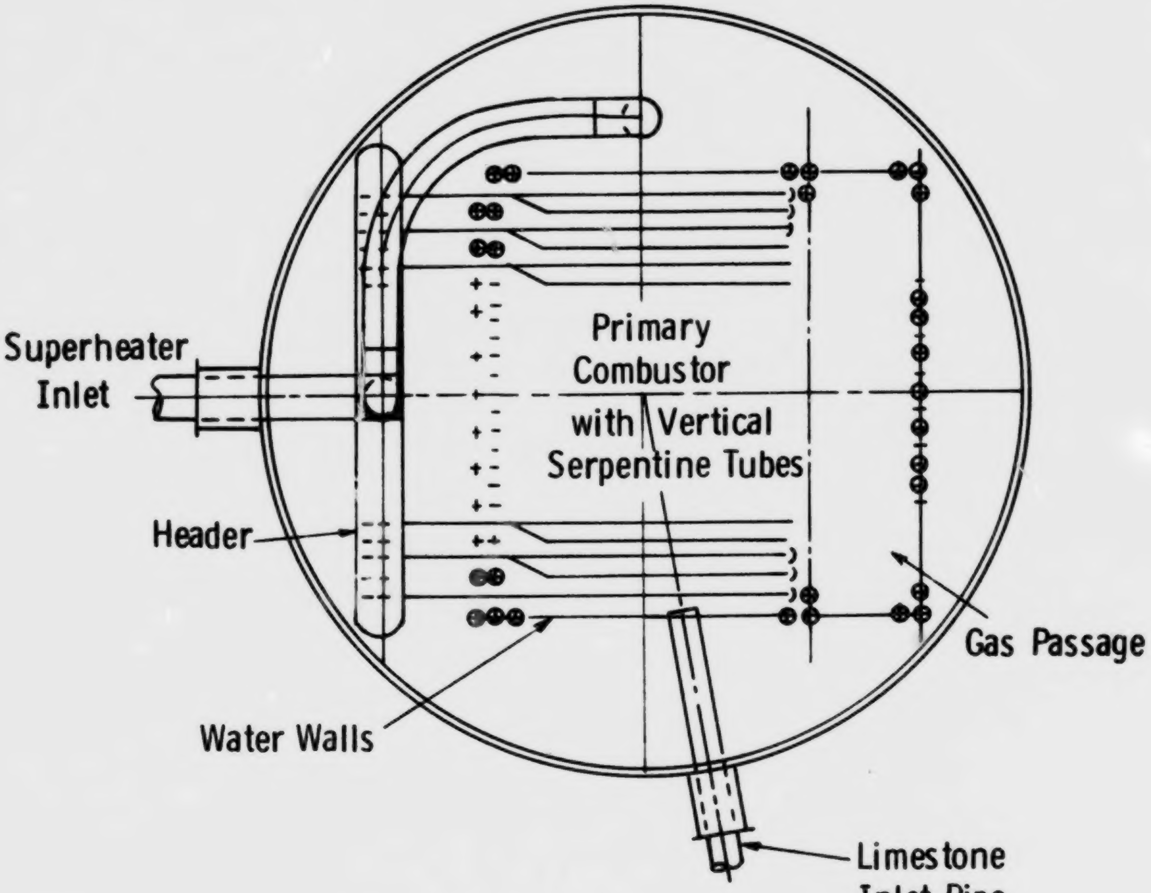
Superheater Inlet
Header
Water Walls
Primary Combustor with Vertical Serpentine Tubes
Gas Passage
Limestone Inlet Pipe

Figure 3-Boiler Cross-Section (Superheater)

TABLE 1

BOILER DESIGN

Design Parameter	Specification
Operating Conditions	
Bed temperature, °F	1300-1750
Gas velocity, fps	5-10
Particle size, inches	1/4" x 0
Excess air, %	10 or 100
Bed depth, ft	10-15
Design Specifications	
Construction	maximum shop fabrication
Heat transfer surface	
Materials	conventional
Tube size, inches	1-1/2 and 2
Gas side heat transfer coefficient, Btu/hr-ft^2-°F	50
Particle carry-over (grains/SCF)	∿ 10
Sulfur removal	
Sorbent	dolomite
System	once-through and regeneration

- Sulfur removal system economics. The operation of the pressurized fluidized bed combustion process using limestone or dolomite results in the production of large quantities of dry, granular partially utilized limestone or dolomite materials in the form of calcium sulfate/calcium carbonate. This sorbent material may be regenerated to a form suitable for repeated sulfur dioxide removal and the sulfur recovered as sulfur or sulfuric acid; or it can be disposed of in its partially utilized form in a once-through system. Alternative regeneration processes and once-through processes were studied, environmental factors identified, economics projected, performance projected, and the processes compared.

- Plant operation and control. The ability of the pressurized fluid bed combustion boiler power plant to start up, shut down, follow load demand, and handle emergencies was conceptually presented in the previous design and evaluation.[1] A more detailed assessment has been made to determine the ability of the concept to meet electrical utility operating requirements. A definition of procedures with estimates of key parameters is presented.

- Alternative concepts. As previously indicated, pressurized fluid bed combustion can be utilized in numerous power cycle concepts and can take on different configurations. Two concepts were studied -- an adiabatic combustor combined cycle plant which represents a modification of the previous power cycle and a recirculating bed boiler which represents an alternative fluidized bed boiler concept. Conceptual designs, performance, and economics are projected and compared with the pressurized fluid bed boiler combined cycle plant.

The results from these studies of pressurized fluid bed combustion boiler combined cycle power plants indicate that:

- Plant costs and performance are essentially invariant with projected changes in operating and design parameters, the costs and performance changes representing less than a 3% change in energy cost.
- A once-through sulfur sorbent process is the most attractive for first generation plants. This conclusion is based on:
  - Data which show that high utilization (1.2 to 1.5 Ca/S molar ratio of the sulfur sorbent can be achieved, which minimizes the quantity of solids for disposal or utilization
  - Data which show that spent stone can be disposed of without further processing
  - Data which show that further development work is required before demonstrating a regenerative system on a commercial plant
  - Improved plant efficiency with once-through operation
  - Reduced complexity/greater reliability with once-through operation
  - Data which show that there are sufficient quantities of sorbent for either once-through or regenerative operation[11]

  The high stone utilization results make the once-through system more economical than previously reported[1], where three to six times the stoichiometric Ca/S ratio was assumed.
- Plant operation and control can meet the requirements of the utility industry; and the techniques proposed for load control, start-up, shutdown, and emergencies are within the state-of-the-art. Plant operation has been developed to provide the capability of operating from 12-1/2% to 100% of rated capacity, change load at the rate of 5% per minute, operate with some components out of service for maintenance, start up or shut down the plant in

several different sequences to ensure maximum usage of plant equipment, and operate the plant as a swing or base load plant.

- Alternative pressurized fluid bed combustion processes should be developed. The adiabatic combustor concept does not offer the performance or economic potential of the pressurized boiler concept. The simplicity, however, and the potential for shorter construction time and for near-term application make the concept attractive. Advanced pressurized boiler concepts such as the recirculating bed boiler offer potential advantages of lower costs, simpler control, and greater sulfur dioxide and nitrogen oxide control. The advanced concepts should be developed as second generation systems.

An assessment of the pressurized fluid bed combustion boiler combined cycle power plant technology based on results from these and other studies show:

1.1 ECONOMIC COMPARISON

Pressurized fluid bed boiler combined cycle plants using state-of-the-art power generation equipment may have energy costs 15 to 20% less than those of conventional plants with stack gas cleaning -- Table 2 and Figure 4. The costs presented are based on the cost data given in the Westinghouse Report of Contract CPA 70-9[1] and EPA data on stack gas cleaning.[9] The costs reported do represent a significant increase from the capital-cost trend of the recent past. However, the costs are considered real for the design basis specified* and are comparable with other contemporary power plant cost estimates. Cost estimates for new systems must be compared with conventional plant costs prepared using the same basis -- thus the relative costs on the same basis are important.

---

*Factors affecting plant costs include the type of plant (e.g. base load versus intermediate load), regional differences (e.g. labor requirements and rates, site conditions), differences in plant facilities (e.g. indoor or outdoor plant, thermal emission control) and differences in what reported estimate includes (e.g. contingency, year of operation, escalation rate).

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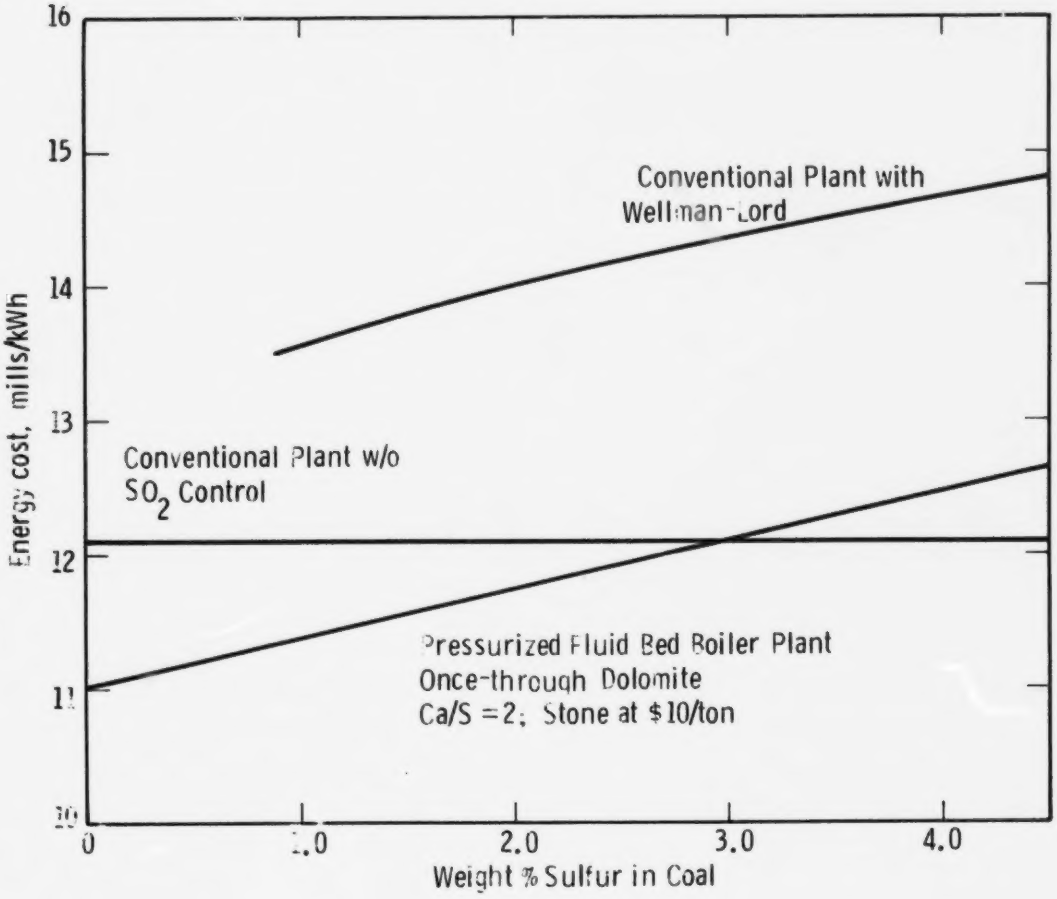


Figure 4–Electrical Energy Costs

TABLE 2

ECONOMIC COMPARISON 600 MW PLANTS[a]
(3% Sulfur Coal)

	Pressurized Fluid Bed Boiler Combined Cycle Once-through Dolomite Sulfur Removal System	Conventional Coal-Fired Plant Wellman-Lord Sulfur Removal System	Conventional Coal-Fired Plant w/o Sulfur Removal System
Plant Capital Cost, $/kw	269	350	306
Energy Costs, mills/kWh			
Fixed charges	6.55	8.60	7.48
O&M	0.71	1.21	0.67
Fuel	4.09	4.55	4.11
	0.75 (dolomite)		
TOTAL	12.10	14.36	12.26

[a] 15%/year
70% capacity factor
No sulfur credit
Coal @ 45¢/10^6 Btu
Methane @ 80¢/10^6 Btu (Wellman-Lord)
Dolomite at $10/ton (purchase plus disposal)
1975 operation of fluid bed boiler plant; 1976 operation of conventional plant.

1.2 ENVIRONMENTAL IMPACT

The ability to meet proposed sulfur dioxide, nitrogen oxide, and particulate emission standards has been demonstrated. Thermal emissions are reduced 15 to 25%. Solids from the plant operated with once-through sorbent include coal ash and dry, granular calcium sulfate/calcium carbonate -- both suitable for disposal without further processing. Spent stone may be regenerable,in which case the quantity of solids would be reduced. Thus, regenerative systems should continue to be studied and developed.

1.3 ENERGY RESOURCES

Fuel conservation is achieved with higher plant efficiencies -- ∿ 10% reduction in fuel requirement with once-through pressurized fluid bed boiler plant using state-of-the-art power generation equipment when compared with a conventional plant with stack gas cleaning. The fluidized bed system also permits the use of low-grade fuels which are not easily used or cannot be used in conventional systems -- e.g. solid wastes, wood chips, refinery bottoms, coal char from gasification systems.

1.4 TECHNICAL UNCERTAINTIES

The economic sensitivity study on operating and design parameters shows that the uncertainties for successful application are minimized, since cost and performance are essentially invariant with the projected design basis. However, back-up design and operating alternatives have been specified where problem areas have been identified. An illustration of the problem areas, proposed solutions, and back-up is presented in Table 3. The design may be conservative in some areas. For example, if deep beds are practical, only one superheat bed may be required; and the carbon burn-up cell,which is desirable in a pressurized system, may not be required with the deep beds and high excess air.

TABLE 3

PROBLEM AREAS

	Proposed Design	Primary Back-up	Alternatives
Particulate Removal[1]	cyclones and Aerodyne-type dust collectors	• provision for third stage • reduce gas velocity • reduce fines content of solids feed • alternative system: granular bed filter	• cool gas and reheat prior to gas turbine • modify turbine • drop back on gas-turbine operating conditions
Sulfur Dioxide Control	dolomite, in bed; 1.2 to 2 Ca/S	• may have to increase use rate for some sorbents	• select new stone
Nitrogen Oxide	minimized during combustion (demonstrated)	• promote reducing conditions in lower region of the bed	
Materials	conventional boiler tube materials	• use higher-grade materials (which are available)	
Coal Feed	petrocarb feed system (lock hoppers)	• increase number of feed points per bed • reduce unit capacity: bed depth	• alternative technology e.g., slurry feed, screw feeders
Alkali Metals[1]	temperature maintained sufficiently low to avoid problem	• lower temperature further to avoid problem	• add sorbent to remove alkali metals • modify turbine operation
Turndown and Load Follow	vary bed temperature modular boilers	• vary excess air	• additional modules • recirculating bed
Spent Stone Disposal	landfill	• sulfur recovery • commercial utilization	

---

[1] Control to achieve gas-turbine reliability and long life

1.5 DEVELOPMENT STATUS

Fluidized bed combustion for the purpose of power generation has come under study, in various forms, in a number of countries during the past twenty or more years. The first major development effort in the United States was initiated in 1965 by the Office of Coal Research (OCR) with Pope, Evans and Robbins to develop an atmospheric-pressure industrial coal-fired packaged boiler.[7] OCR has a continuing effort to develop an atmospheric-pressure system and is supporting pressurized system development. EPA became involved in the development of fluidized bed combustion to achieve reduced air pollutant emissions in 1967. An overall program has been under way by EPA to demonstrate the technology on a full scale. The program, which has included eight contractors and subcontractors, including Westinghouse; Pope, Evans and Robbins; Esso*; Erie City; Foster Wheeler; Argonne National Laboratories; the Bureau of Mines; and the National Coal Board, represents a broad spectrum of capabilities. Pope, Evans and Robbins, under contract to OCR, is designing a 30 MW atmospheric-pressure fluid bed boiler. Under the direction of the Office of Research and Development, EPA, pressurized fluid bed combustion units have been built by Argonne National Laboratory and Esso Research and Engineering. A 1 MW test facility is being built by Esso; and a 30 MW development plant has been designed by Westinghouse. The British Coal Utilization Research Association was the first to build and operate a pressurized fluid bed combustor of approximately 1 MW equivalent capacity.[4,10] The pressurized fluid bed boiler development status is summarized in Table 4.

A demonstration plant program has been conceived under the direction of EPA to commercialize the pressurized fluid bed boiler concept. The 30 MW development plant will provide the capability to study and demonstrate pressurized fluid bed combustion and heat transfer, steam generation, solids feed and handling, particulate removal, gas-turbine performance, boiler control, and sulfur dioxide and nitrogen oxide emission control. Preliminary designs, cost estimates, experimental program, schedule, and program alternatives are presented in Volume III. A model of the plant, estimated to cost approximately $10 million, is shown in Figure 5.

---

*Esso has since changed its name to Exxon. For purposes of this report, the name Esso has been retained.

TABLE 4

PRESSURIZED FLUID BED BOILER DEVELOPMENT STATUS

Apparatus

- Pressurized combustion, sulfur removal and regeneration test units operating ($\leq$ 1 MW equivalent)
- Pressurized fluid bed boiler pilot plant operating ($\sim$ 1 MW equivalent)
- Second pressurized fluid bed boiler pilot plant to be operational in 1974 ($\sim$ 1 MW equivalent)

Designs

- Preliminary design of commercial plant complete (635 MW)
- Demonstration boiler plant preliminary design complete (20-30 MW equivalent)

Results

- 95% sulfur removal demonstrated
- $<$ 0.2 lb $NO_2/10^6$ Btu demonstrated
- No erosion, deposition, corrosion of gas-turbine blades after 200 hr tests
- Boiler tube materials adequate for commercial operation
- Horizontal tubes tested
- Coal feeding at pressure demonstrated (5 atm; 10 atm tests under way)
- Continuous operation - runs up to 350 hours
- Economic sensitivity of design and operating conditions checked

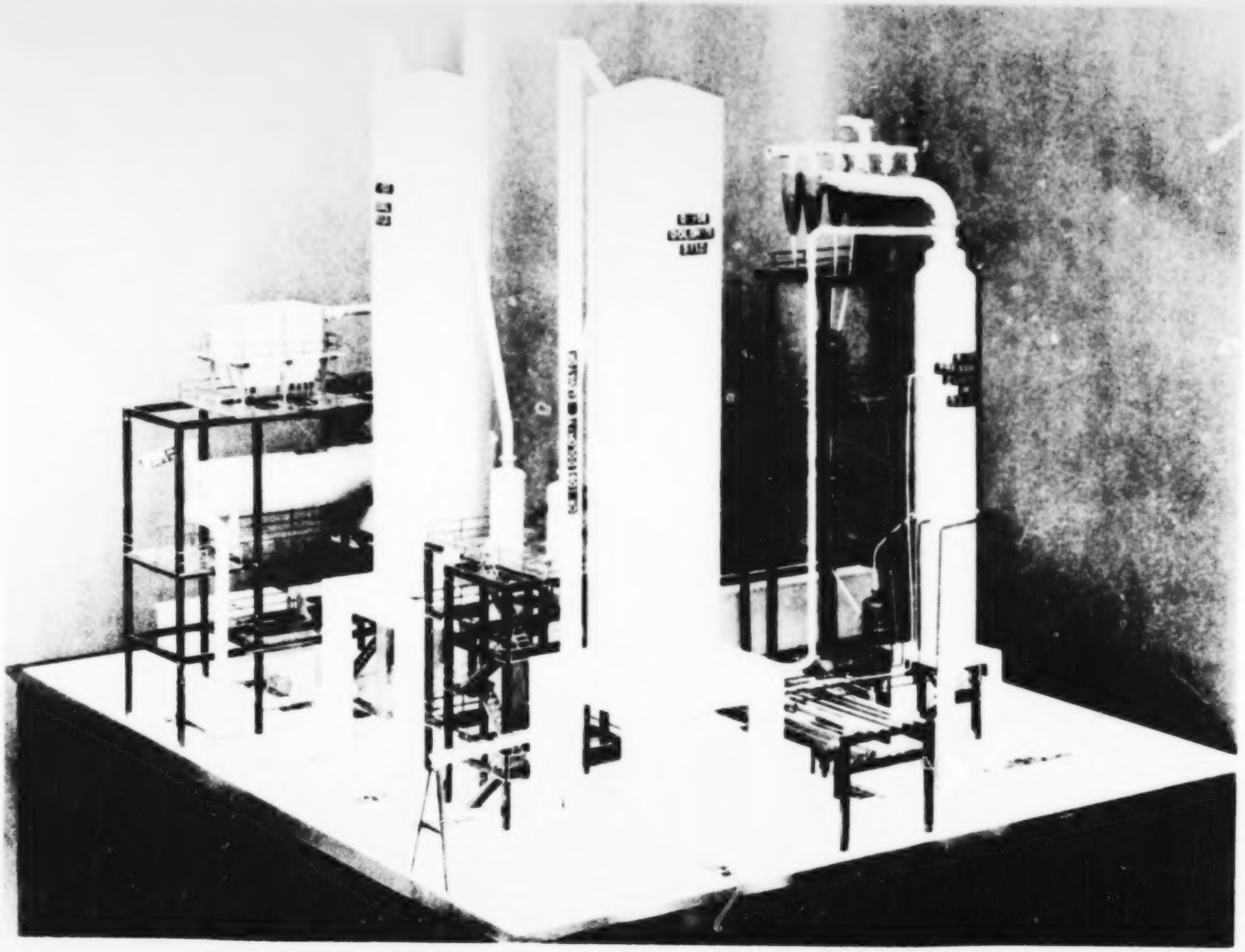


Figure 5 30 MW Pressurized Fluid Bed Boiler Demonstration Facility

1.6 COMPARISON WITH ALTERNATIVE TECHNOLOGY

The pressurized fluid bed combustion boiler concept using state-of-the-art power generation equipment is projected to meet environmental requirements and to have energy costs as low as or lower than any other competitive system -- commercial or under development -- using state-of-the-art power generation equipment. This includes stack gas cleaning, liquefaction/power generation, gasification, atmospheric-pressure fluidized bed combustion, or low-sulfur fuel. The concept also offers greater simplicity and versatility than competitive systems using high-sulfur/low-grade fuels by utilizing only one fuel processing step -- combustion -- and by controlling sulfur dioxide and nitrogen oxide emissions during the fuel processing step. Thus, the system avoids fuel pretreatment and processing of the stack gas. The system can be demonstrated for use in the 1980s. Second generation pressurized fluid bed combustion boiler systems offer the potential for increased plant efficiency (up to ∿ 45%) by increased gas-turbine temperature and increased steam temperature and pressure[1,2,8] and the potential for application in alternative and advanced power cycles.[3,5] In conclusion:

- Pressurized fluid bed combustion combined cycle power plants using commercial power generation equipment offer the opportunity to generate electrical energy within environmental constraints at lower energy costs than competitive systems.
- No problems have been identified which preclude the development of pressurized fluid bed combustion combined cycle power plants.
- Demonstration of the pressurized fluid bed combustion combined cycle power plant concept should be accelerated; detailed design and construction of a 30 MW demonstration facility should begin.

2. ECONOMIC SENSITIVITY

A pressurized fluidized bed boiler power plant has been designed using state-of-the-art power generation equipment,[1,2] with performance, costs, and pollution abatement projected for the system. The results show the concept has the potential to meet SO_2, NO_x, and particulate emission standards; and could reduce energy costs 10% below those of a conventional plant with stack gas scrubbing.

Operating conditions and design parameters for the pressurized boiler were selected on the basis of an evaluation of available data, power cycles, and alternative boiler concepts. It is important to know how sensitive the operating and design parameters selected for the base design are to the plant economics. An understanding of the effect of changes in the proposed design on plant cost will provide a basis for evaluating current pressurized fluid bed combustion pilot plant data, planning experimental programs, designing the development plant, and understanding the economic margin for solving technological problems.

The sensitivity analysis evaluates the effect of the following variables on plant design, cost, and performance:

- Fluidized bed boiler operating conditions
  - Bed temperature
  - Fluidizing velocity
  - Excess air
  - Pressure
- Fluidized bed boiler design
  - Heat transfer surface: configuration, heat transfer coefficient, materials
  - Module capacity
- Particulate carry-over from the boiler
  - Loading
  - Size distribution

- Power plant equipment operating conditions
  - Gas-turbine inlet temperature
  - Steam temperature and pressure.

The evaluation is performed by considering each variable separately. It is general rather than specific in order to permit the coupling of different effects and thereby assess alternative designs; and it is performed for the purpose of indicating the effects of variable changes in plant costs relative to each other and to the total plant cost.

2.1 BASIS FOR SENSITIVITY ANALYSIS

The basis for the sensitivity analysis is the boiler and plant design developed by Westinghouse under contract to EPA.[1,2] The power plant cycle is shown schematically in Figure 6. The plant subsystems included in this sensitivity analysis are enclosed within the broken lines. The pressurized boiler designed by Westinghouse and Foster Wheeler is shown schematically in Figure 7. The preliminary boiler design was for a nominal 300 MW plant. The boiler design consists of four modules and provides for a maximum of shop fabrication and turndown requirements. Each module includes four primary fluidized bed combustors, each containing a separate boiler function -- one bed for the preevaporator, two beds for the superheater, and one bed for the reheater. Evaporation takes place in the water walls. All of the boiler heat transfer surface is immersed in the beds, except for baffle tubes above the bed to minimize particle carry-over. Each module contains a separate fluidized bed or carbon burn-up cell to complete the combustion of carbon elutriated from the primary beds. The philosophy behind the boiler design was the maximization of shop fabrication. Thus, the 300 MW plant utilizes boiler modules which can be completely shop fabricated. From roughly 300 to 600 MW, the boiler modules can be only partially shop fabricated since the pressure shell is too large for rail transport. Larger plants utilizing the four-module concept would be field erected.

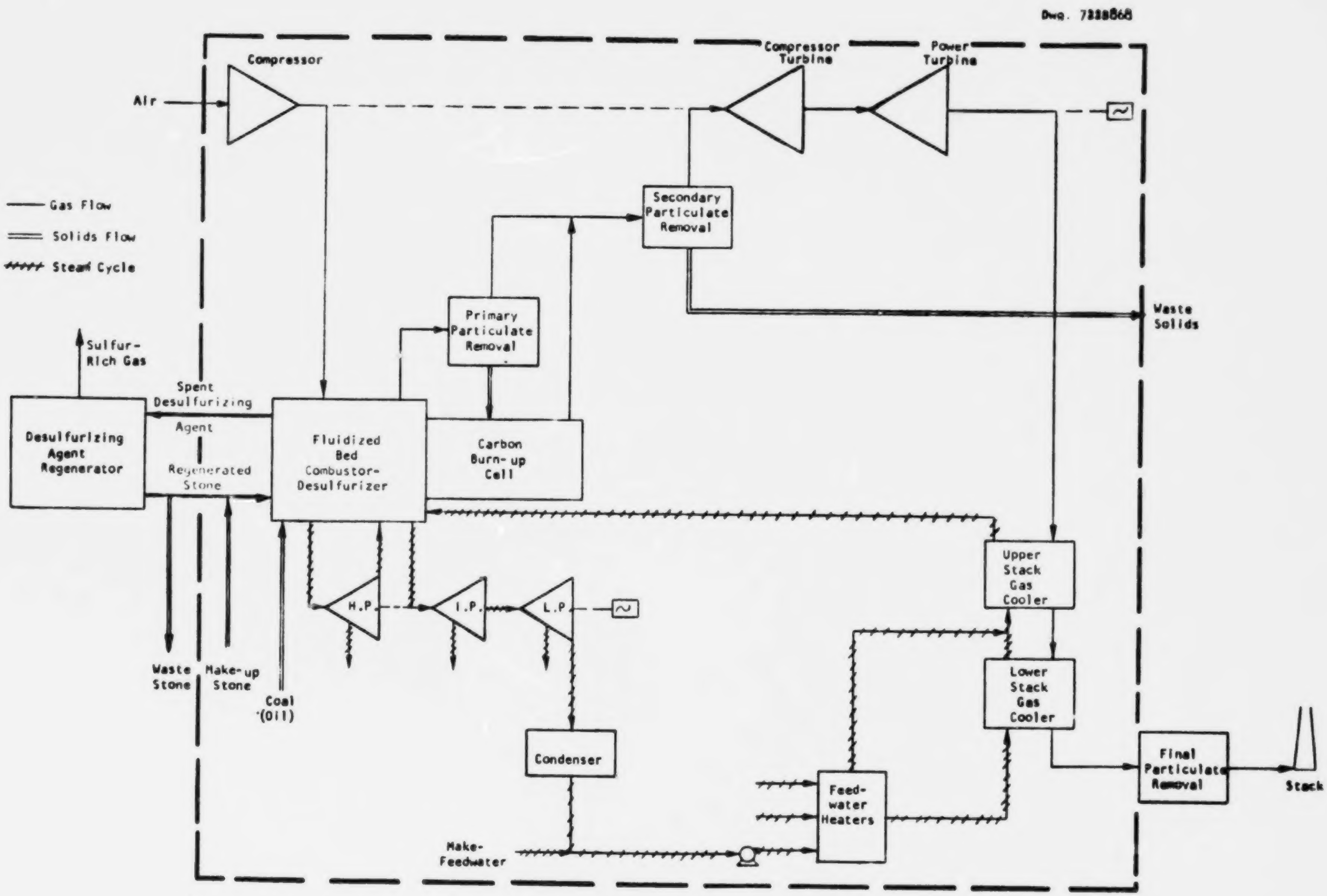
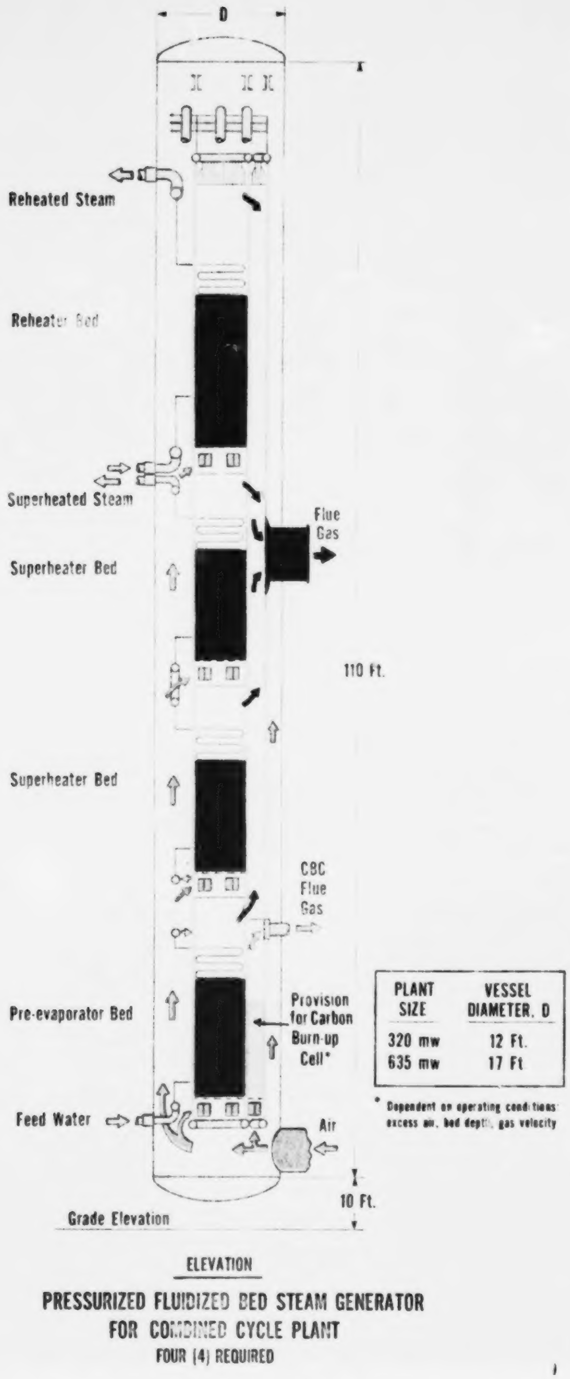

Dwg. 7338868
Compressor
Air
Compressor Turbine
Power Turbine
~
Gas Flow
Solids Flow
Steam Cycle
Secondary Particulate Removal
Primary Particulate Removal
Waste Solids
Sulfur-Rich Gas
Spent Desulfurizing Agent
Desulfurizing Agent Regenerator
Regenerated Stone
Fluidized Bed Combustor-Desulfurizer
Carbon Burn-up Cell
Upper Stack Gas Cooler
H.P.
I.P.
L.P.
~
Waste Stone
Make-up Stone
Coal (Oil)
Lower Stack Gas Cooler
Condenser
Final Particulate Removal
Stack
Feed-water Heaters
Make-Feedwater

Figure 6: Pressurized Fluid Bed Boiler Power System


D
Reheated Steam
Reheater Bed
Superheated Steam
Flue Gas
Superheater Bed
110 Ft.
Superheater Bed
CBC Flue Gas
Pre-evaporator Bed
Provision for Carbon Burn-up Cell*
Feed Water
Air
10 Ft.
Grade Elevation
ELEVATION
PRESSURIZED FLUIDIZED BED STEAM GENERATOR
FOR COMBINED CYCLE PLANT
FOUR (4) REQUIRED

PLANT SIZE	VESSEL DIAMETER, D
320 mw	12 Ft.
635 mw	17 Ft

* Dependent on operating conditions: excess air, bed depth, gas velocity

Figure 7: Pressurized Fluidized Bed Steam Generator Module

The operating conditions and design parameters for the boiler and the power cycle are summarized in Table 5. The power plant performance and economics were determined on the basis of these specifications. The cost breakdown for the fluidized bed steam generator is summarized in Table 6. The fluidized bed boiler design was scaled to 600 MW capacity and the costs estimated. These costs are summarized in Figure 8. A breakdown of the power plant equipment costs for a 635 MW plant is presented in Table 7. Limestone or dolomite regeneration is not included in this analysis: the costs presented are for a once-through system. The energy costs used for this analysis are also presented in Table 8. The costs of a conventional plant with wet scrubbing on the same basis are also indicated.

The following assumptions are made for the sensitivity analysis:

- The plant maintains the four-module concept with two modules per gas-turbine.
- The coal feed rate is maintained constant for each variable analysis. Thus, the coal feeding and handling system is assumed to remain unchanged. This may not be completely true if the bed area is changed significantly and the number of feed points increased or decreased. The cost of the coal feed system is considered if the bed design is altered.
- Structural and erection costs are constant. The structural steel and concrete costs for the boiler plant equipment are ∿ \$2/kw. The maximum change in these costs for the cases considered is ∿ \$0.20/kw and, in general, will be significantly less. The cost is thus assumed constant for this analysis. Erection cost changes are negligible.

TABLE 5

PRESSURIZED FLUIDIZED BED BOILER POWER PLANT OPERATING AND DESIGN CONDITIONS

Item	Specification
Cycle	
Steam system	2400 psia, 1000°F superheat, 1000°F reheat
Gas-turbine expander	
Pressure ratio	10:1
Inlet temperature	1600°F
Air cooling	5%
Coal feed rate	53,910 lb/hr/module for nominal 300 MW plant design
Number of boiler modules	4
Boiler modules/gas turbine	2
Fuel/air ratio	0.0919
Boiler Design	
Bed area	35 ft^2 (5' x 7') -- for ∿ 80 MW module
Heat transfer surface	
Walls	2" O.D. tubes on 3-1/2" welded wall spacing
Bed	1-1/2" O.D. tubes in preevaporator and and superheater; 2" O.D. in reheater (details in text)
Gas side heat transfer coefficient	50 Btu/hr-ft^2-°F
Tube materials	SA-210-A1 -- preevaporator SA-213-T2 -- lower superheater SA-213-T22 -- water walls; upper superheater (lower loops); reheater SA-213-TP304H -- upper superheater (upper loops)
Bed depth (expanded)	11 to 14 ft
Gas temperature drop from primary beds to gas-turbine expander	150°F

TABLE 5 (Continued)

Item	Specification
Boiler Operating Conditions	
Bed temperature (100% load)	1750°F
Fluidizing velocity	6 to 9 fps
Excess air	17.5%
Particle carry-over	∿ 7 grains/SCF
Carbon from primary beds	6% of carbon feed
Auxiliaries	
Coal feed system	petrocarb feed system
Primary particulate removal	4 size 355 VM 8/0/150 Duclone per module -- nominal 300 MW design
Secondary particulate removal	2 model 18000 Type S collectors per module -- nominal 300 MW design (quoted by Aerodyne Dev. Corp.)
Stack gas coolers	conventional heat exchanger design

TABLE 6

COST OF A 318 MW PRESSURIZED FLUID BED BOILER

Item	Cost
Pressure Parts[a]	$1,777,000
Shell	935,000
Subcontracted & Contracted Equipment	435,000
Drafting	185,000
Home Office	685,000
Subtotal	$4,017,000 field-erected
	($3,856,000 shop-assembled)
Erection	500,000
TOTAL	$4,517,000

[a] Pressure parts include tubing cost, headers, downcomers, risers, tube bending, tube welding, and water-wall fabrication.

TABLE 7

EQUIPMENT COSTS FOR PRESSURIZED FLUIDIZED BED BOILER POWER PLANT[1] (635 MW)

Component	Cost($/kw)
Boiler Plant Equipment	
Boiler	14.49
Particulate removal	12.76
Piping/ducts	4.43
Stack and foundation	0.47
Coal handling and feeding equipment	14.94
Ash and dust handling system	1.55
Instruments and controls	3.10
Miscellaneous equipment	0.94
Steam Turbine - Generator Equipment	44.14
Gas Turbine - Generator Equipment	14.80
Other (land, structures, electric plant equipment, miscellaneous plant equipment, undistributed costs)	70.38
SUBTOTAL	182.00
TOTAL CAPITAL COST (inc. escalation, IDC, etc.)	265.00

($340/kw for conventional plant with wet scrubbing on same basis)

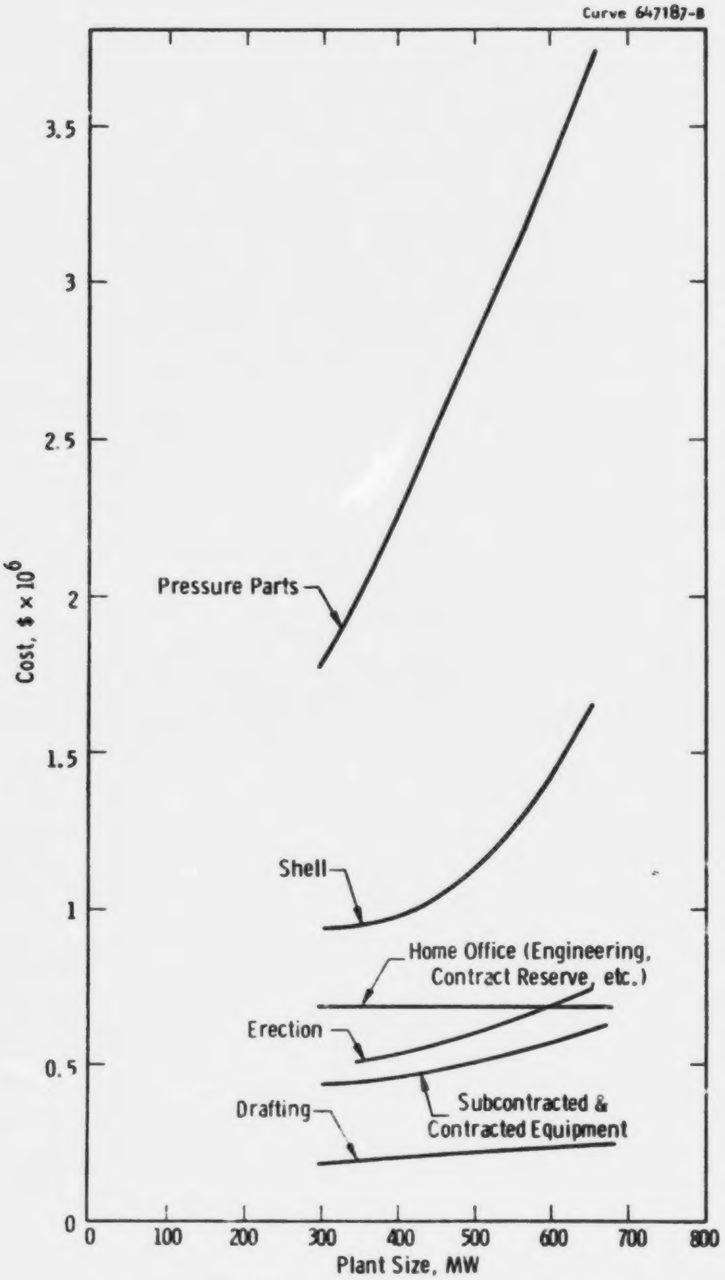


Figure 8: Steam Generator Cost Breakdown for the W-FW Basic Design

TABLE 8

ENERGY COSTS FOR PRESSURIZED FLUIDIZED BED BOILER

Item	Cost (mills/kWh)
Fixed Charges	6.44
Fuel	4.04
Dolomite	0.52
Operating and Maintenance	0.71
	11.71

(13.45 for conventional plant)

- Coal and stone feed size are assumed constant: 1/4" x 0. This parameter is important when considering particle carry-over, but insufficient information is available to permit a quantitative analysis.

2.2 OPERATING CONDITIONS

2.2.1 Bed Temperature

The full load design temperature is 1750°F. Lowering the design bed temperature increases the total amount of heat transferred in the bed and thus increases the steam-turbine power generation. At the same time, lowering the bed temperature decreases the gas-turbine inlet temperature -- assuming there is no burning above the bed -- and thus decreases the total gas-turbine power generation. The decrease in gas-turbine power is larger than the increase in steam power and results in an overall decrease in plant power (Figure 9). Lowering the bed temperature increases the total amount of heat transferred in the bed and thus requires more heat transfer surface (Figure 10). Assuming the cross-sectional area of the fluid bed (5 ft x 7 ft for a 300 MW nominal plant size) and tube size/tube pitch are constant, the expanded bed depth for each functional bed increases with a decrease in bed temperature, as shown in Figure 11. The bed depth of the two superheater beds is assumed for convenience to be the same. This will not affect the total heat transfer surface requirement and the resultant module height shown in Figure 12. The bed depth and module height can be reduced by enlarging the bed area and module diameter. However, since the module diameter of 12 feet is considered to be the largest size that can be shipped by railroad, an increase in the module diameter to accommodate additional heat transfer surface may not be economical for a 300 MW plant.

The effect of changing design bed temperature on the steam generator cost is calculated on the basis of the boiler cost estimation shown in Table 6 and on the assumptions that the module diameter is constant

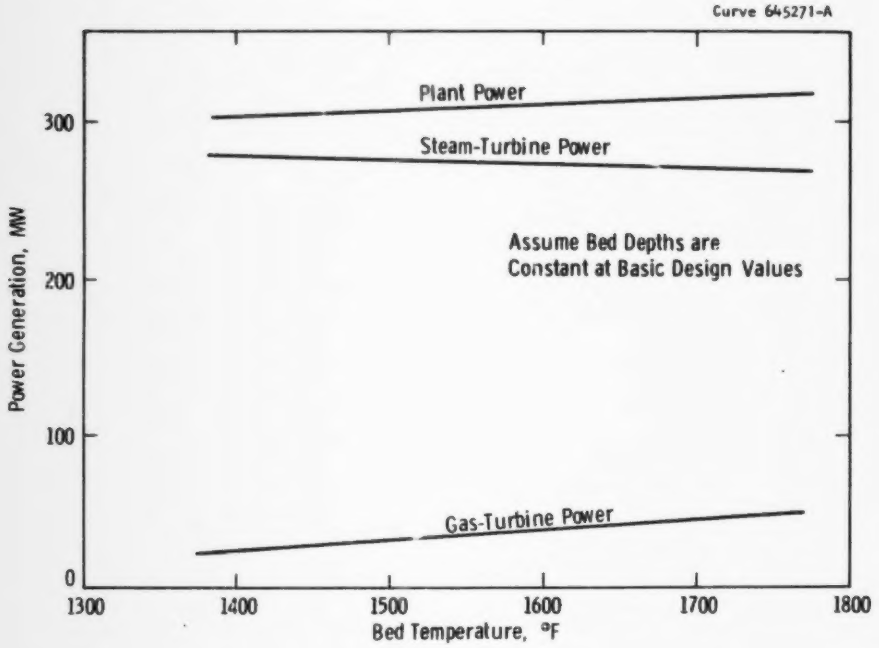


Figure 9: Effect of Bed Temperature on Power Generation

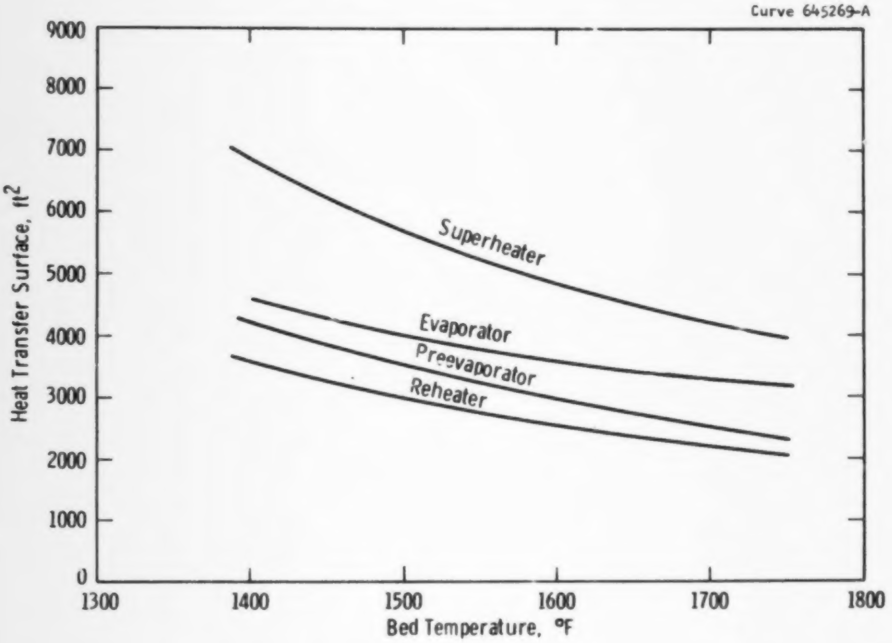


Figure 10: Heat Transfer Surface Requirement for One Module

Curve 645270-A

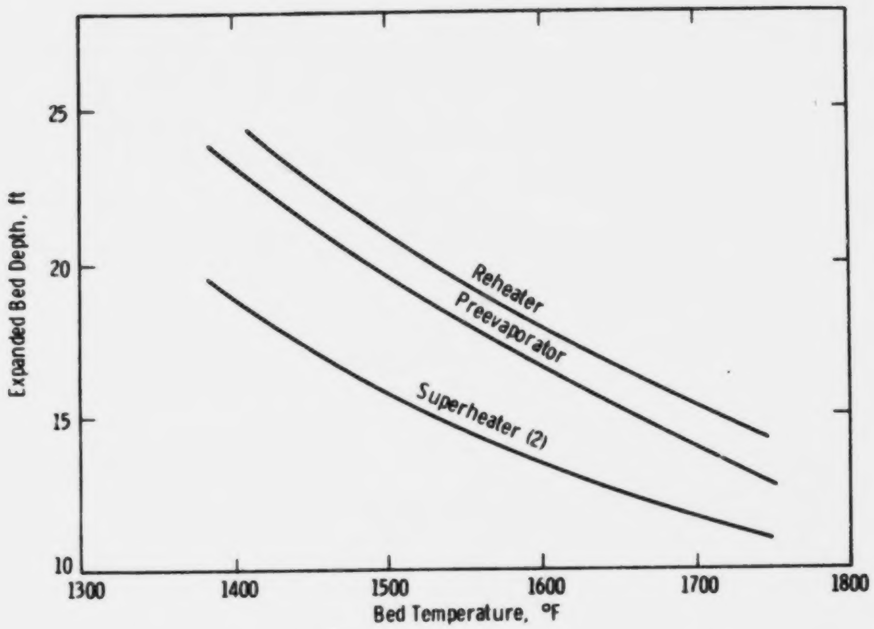


Figure 11: Bed Depth Requirement at Different Bed Temperatures

Curve 645273-A

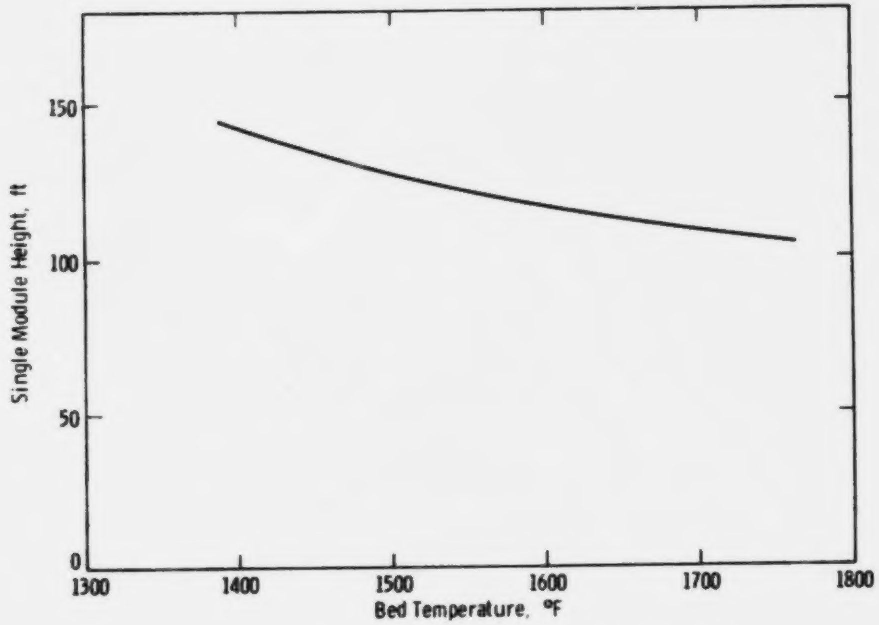
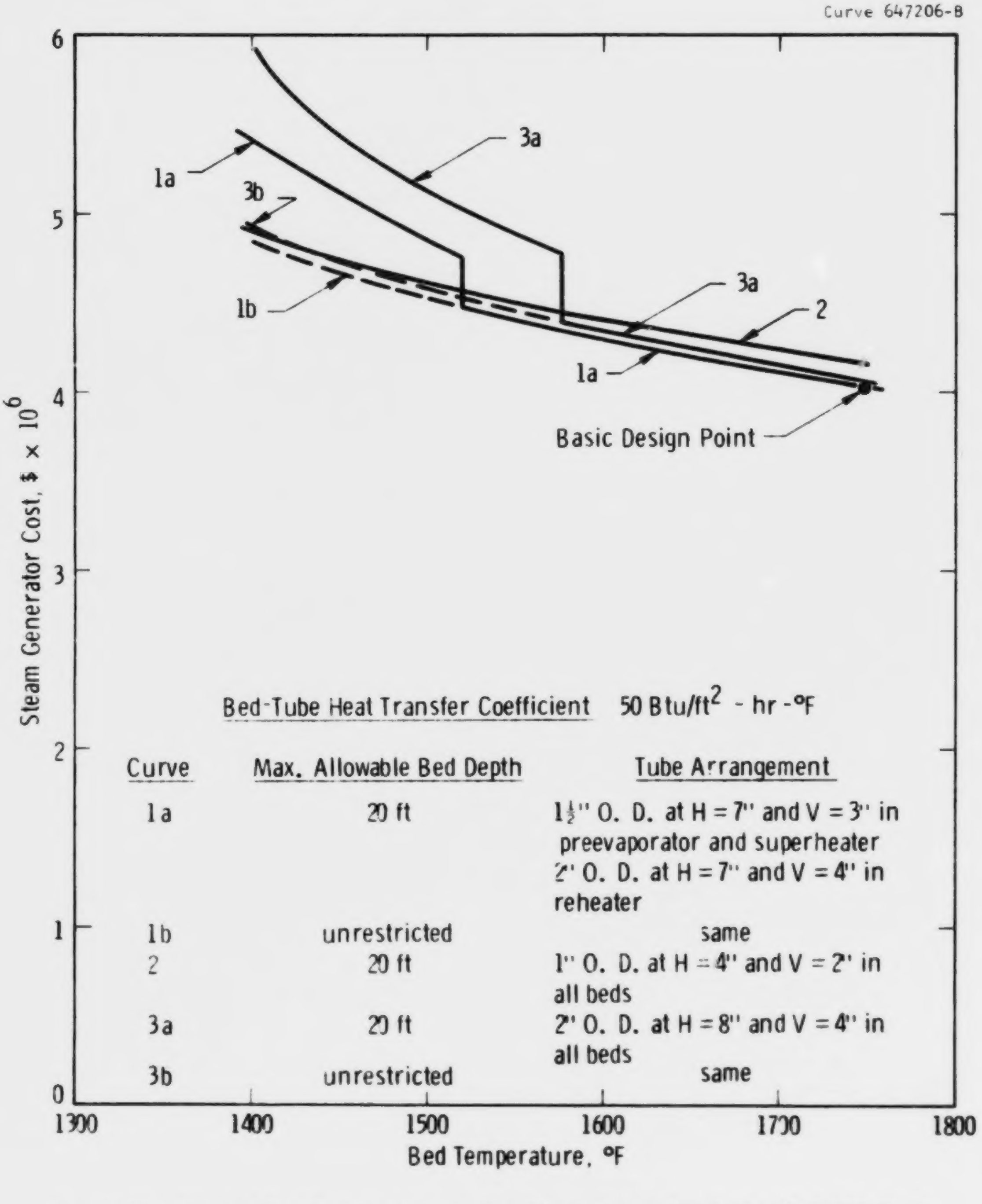


Figure 12: Effect of Bed Temperature on Total Module Height

at 12 ft and the total number of modules is four, based on turndown consideration. The cost (not including erection) of the four-module steam generator as a function of bed temperature with constant module diameter (12 ft) is shown as curves 1a and 1b in Figure 13. Curve 1a assumes the maximum allowable bed depth to be 20 ft. That means any bed with an expanded bed depth larger than 20 ft will have to be split into two beds with separate air plenums and freeboards. Curve 1b assumes that there is no restriction on maximum bed depth. The choice of 20 ft as the maximum allowable bed depth was arbitrary and was just to show the effect of this variable on the cost of a steam generator. It is doubtful that the bed depth of each fluid bed can be unrestricted without creating undesirable bubble formation and slugging, poor bed-tube heat transfer coefficient, and temperature gradients in the bed at some bed depth. The maximum allowable bed depth at specific operating conditions will have to be experimentally determined in a large unit. Without the required experimental evidence, the steam generator cost (not including erection) is plotted against the maximum allowable bed depth in Figure 14. The bed temperatures were calculated by assuming the gas-turbine temperatures of 1600°F, 1500°F, 1400°F, and 1300°F; and by assuming a linear temperature loss between the boiler and the gas-turbine inlet. The cost of the steam generator designed for 1636°F increases ∿ 20% (∿ \$2.8/kw) over that designed at 1750°F if the maximum allowable bed depth is 15 ft. The cost increase is primarily due to the splitting of beds with a bed depth greater than 15 ft. This bed splitting can be avoided either by decreasing the boiler tube diameter and the spacing in the bed or by increasing the module diameter. Decreasing the tube diameter and spacing will change the bed-tube heat transfer coefficient, tube bending and fabrication, and tube-wall thickness; increasing the module diameter will not only change the cost of the pressure shell but will also affect construction -- complete shop assemblage versus partial field erection. All these factors which have to be taken into account in designing an optimal boiler are discussed in separate sections.



Bed-Tube Heat Transfer Coefficient 50 Btu/ft^2 - hr - °F

Curve	Max. Allowable Bed Depth	Tube Arrangement
1a	20 ft	$1\frac{1}{2}$" O. D. at H = 7" and V = 3" in preevaporator and superheater 2" O. D. at H = 7" and V = 4" in reheater
1b	unrestricted	same
2	20 ft	1" O. D. at H = 4" and V = 2" in all beds
3a	20 ft	2" O. D. at H = 8" and V = 4" in all beds
3b	unrestricted	same

Figure 13: Effect of Bed Temperature on the Steam Generator Cost of 318 MW Plant (not including erection)

Curve 647155-A

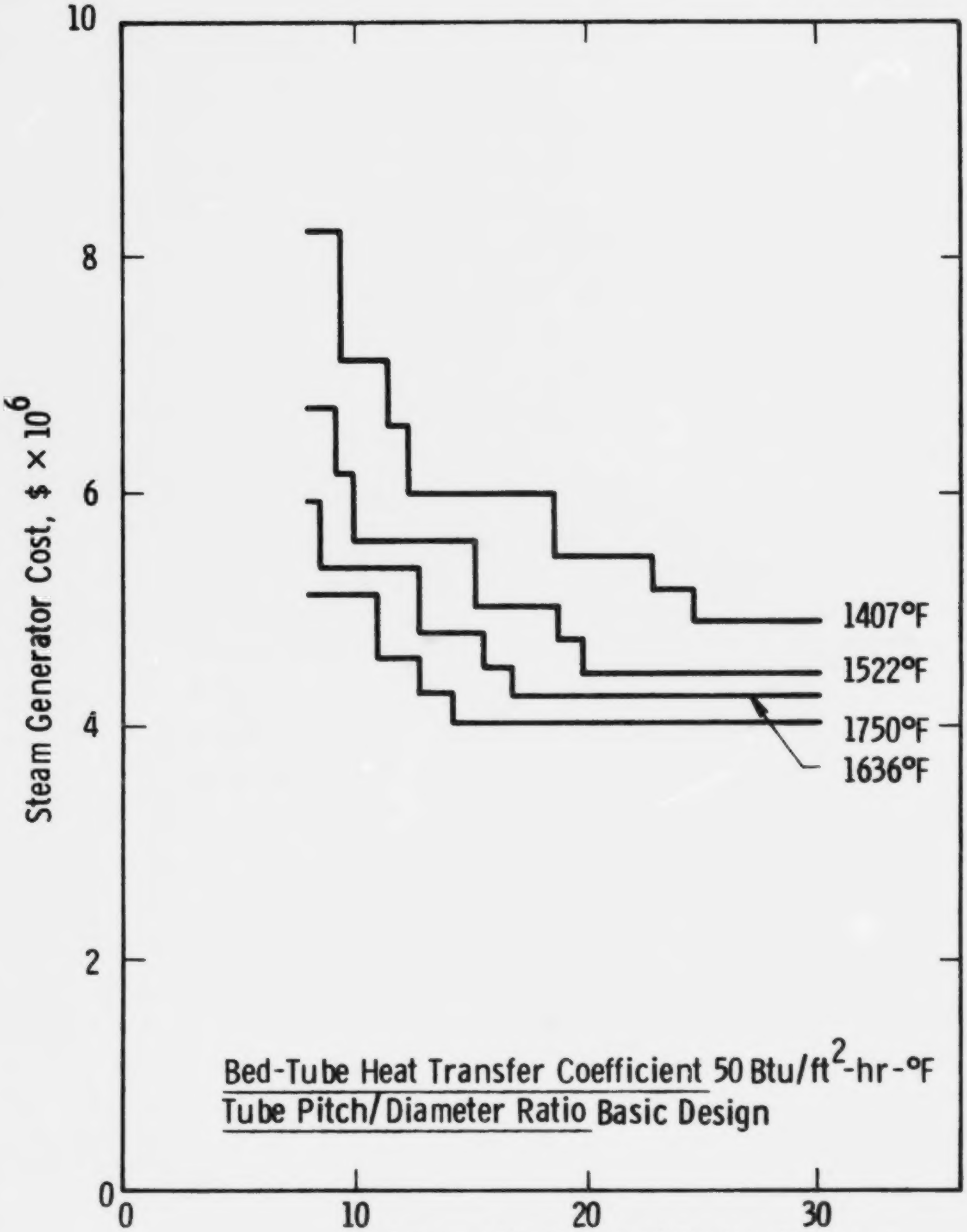


Figure 14: Dependence of the Steam Generator (318 MW) Cost on the Maximum Allowable Bed Depth (not including erection)

In the present cost estimation, the possibility of using thinner-wall tubes for the designs at lower bed temperatures was also taken into account by calculating the minimum tube-wall thickness requirement. No allowance for corrosion was provided. The method for calculating the minimum tube-wall thickness and the results of the calculation are presented in Volume II, Appendix A.

A change in the operating bed temperature will also change the gas temperature to both the primary and secondary cyclones and, thus, change the actual volumetric gas flow rate. This will change the gas inlet velocity to the cyclones which, in turn, affects cyclone collection efficiency. This effect was estimated to be small compared to the effect of the change in pressure drop across the bed due to a change in the design bed temperature. Decreasing the design bed temperature increases the heat transfer surface requirement in the bed. This requires an increase in pressure drop due to an increase in bed depth if the bed area and boiler tube configuration are constant. The decrease in net plant power output as a function of the design bed temperatures is presented in curve 1, Figure 15. The effect is small -- a decrease of only ∿ 0.3% if the design bed temperature is reduced to 1407°F.

Bed temperature is one of the primary variables used for load turndown in the present design. A 4:1 turndown can be met if the design bed temperature is higher than 1600°F. The primary limitation on the operating bed temperature is the sulfur removal efficiency of the sorbents in the bed. At bed temperatures higher than 1750°F or lower than 1350°F, the sulfur removal efficiency in the bed is too low. Thus, it is concluded from the above bed temperature analysis that the design bed temperature should be the highest temperature required for a desirable degree of load turndown and sulfur removal in the bed.

2.2.2 Fluidizing Velocity

At a constant fuel feed rate and excess air, increasing the fluidizing velocity will require a decrease in the bed area and in the

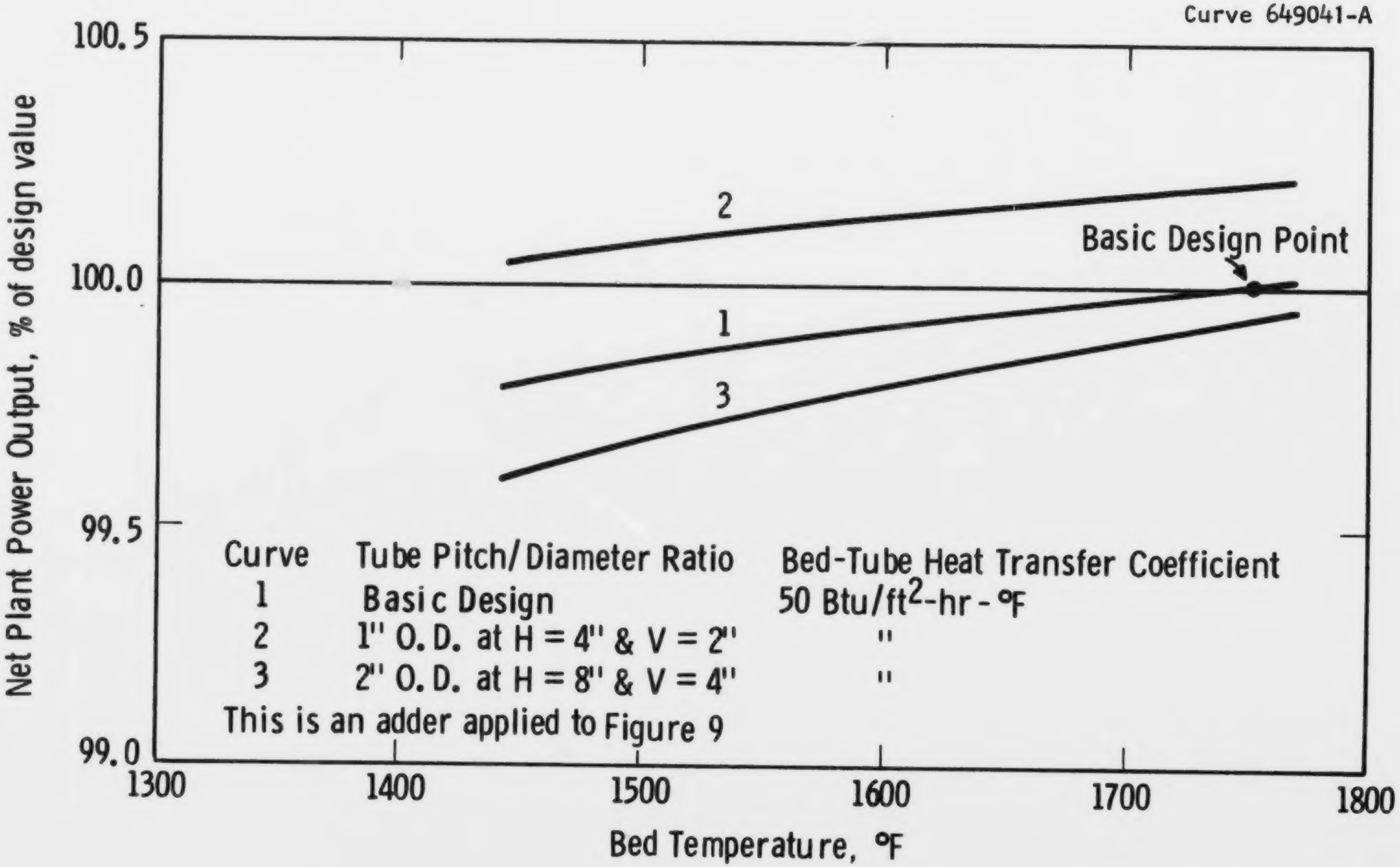


Figure 15: Change in Net Plant Power Output Due to Change in Pressure Drop across the Bed

module diameter. Since, for a constant overall heat transfer coefficient and a specific design bed temperature, the total heat transfer surface in the bed is constant, a decrease in the bed area will require an increase in the bed depth at constant tube size and tube spacing and, thus, an increase in the module height. An economical design will depend on the balance of these factors.

The bed area and bed depth requirements with respect to change in fluidizing velocity at different design bed temperatures were calculated and are shown in Figures 16 through 19. The corresponding module height and module diameter are presented in Figure 20. The cost of the pressure shell at different inside diameters is estimated on the basis of data from Foster Wheeler Corporation[1] and an independent estimation by Westinghouse. This is shown in Figure 21. The discontinuity at a module inside diameter of 12 ft is due to the cost difference between the shop-assembled and the field-erected shell.

Additional fabrication costs must be considered for the shell and for the concomitant change of water walls and tube bending. Taking into account all the factors involved, the steam generator cost is plotted against the superficial fluidizing velocity in the preevaporator as shown in Figure 22 for a 318 MW and a 635 MW plant. The superficial fluidizing velocity in the preevaporator was used here since it is the largest velocity in all beds inside a single module. The superficial fluidizing velocity in the superheaters and reheater can be calculated accordingly. The results show that increasing the fluidizing velocity tends to increase rather than decrease the total steam generator cost of a 318 MW plant. Decreasing the fluidizing velocity in the preevaporator below $\sim$ 8 ft/sec will require a shift from shop assemblage to field erection and will escalate sharply the steam generator cost. At a plant size of 635 MW, a minimum cost does exist. Figure 22 shows costs for unrestricted maximum allowable bed depth. If, for example, the maximum allowable bed depth is limited to 10 ft or 20 ft, the disadvantage of increasing the fluidizing velocity would be even larger at the 318 MW size. At the 635 MW size, the minimum would shift to lower velocity.

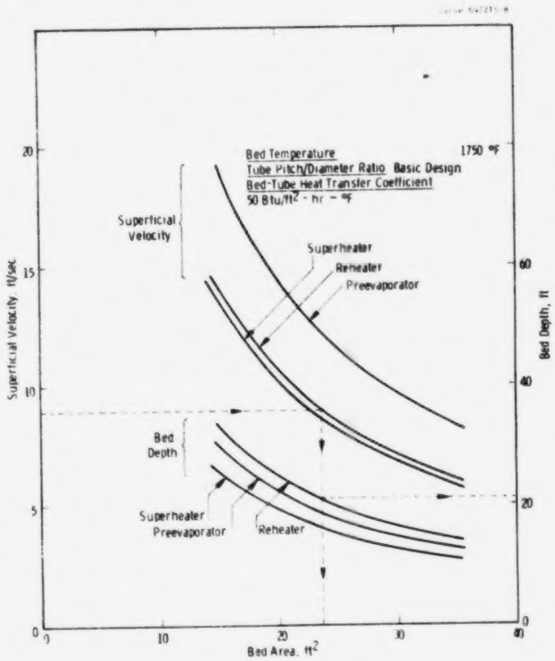


Figure 16 Dependence of Bed Depth on the Fluidizing Velocity. Bed Temperature, 1750°F

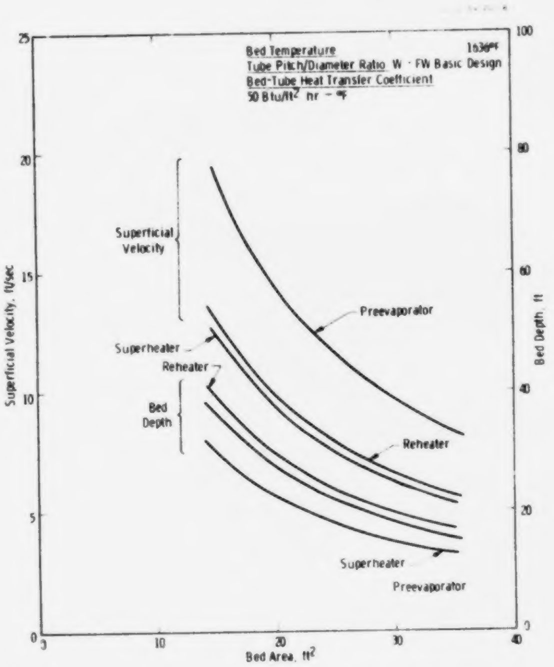


Figure 17 Dependence of Bed Depth on the Fluidizing Velocity. Bed Temperature, 1636°F

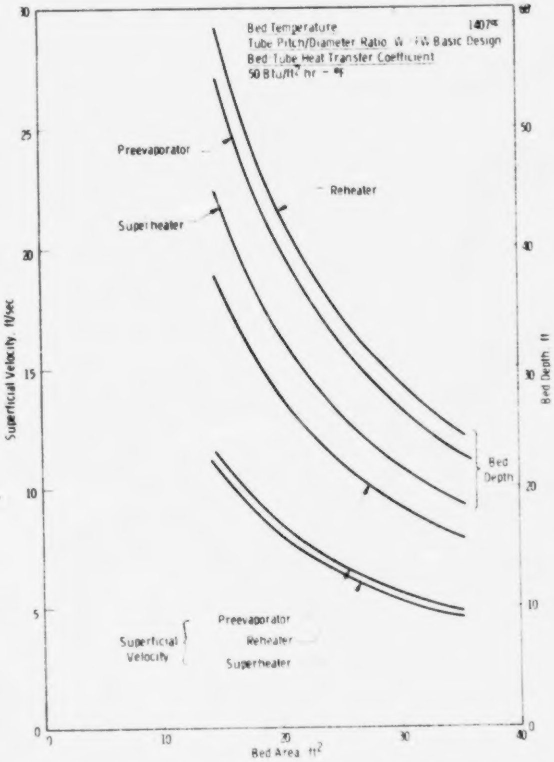


Figure 19 Dependence of Bed Depth on the Fluidizing velocity. Bed Temperature 1407°F

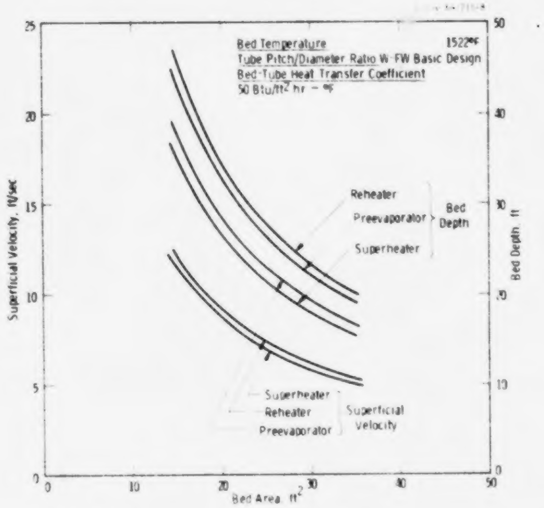


Figure 18 Dependence of Bed Depth on the Fluidizing Velocity. Bed Temperature, 1522°F

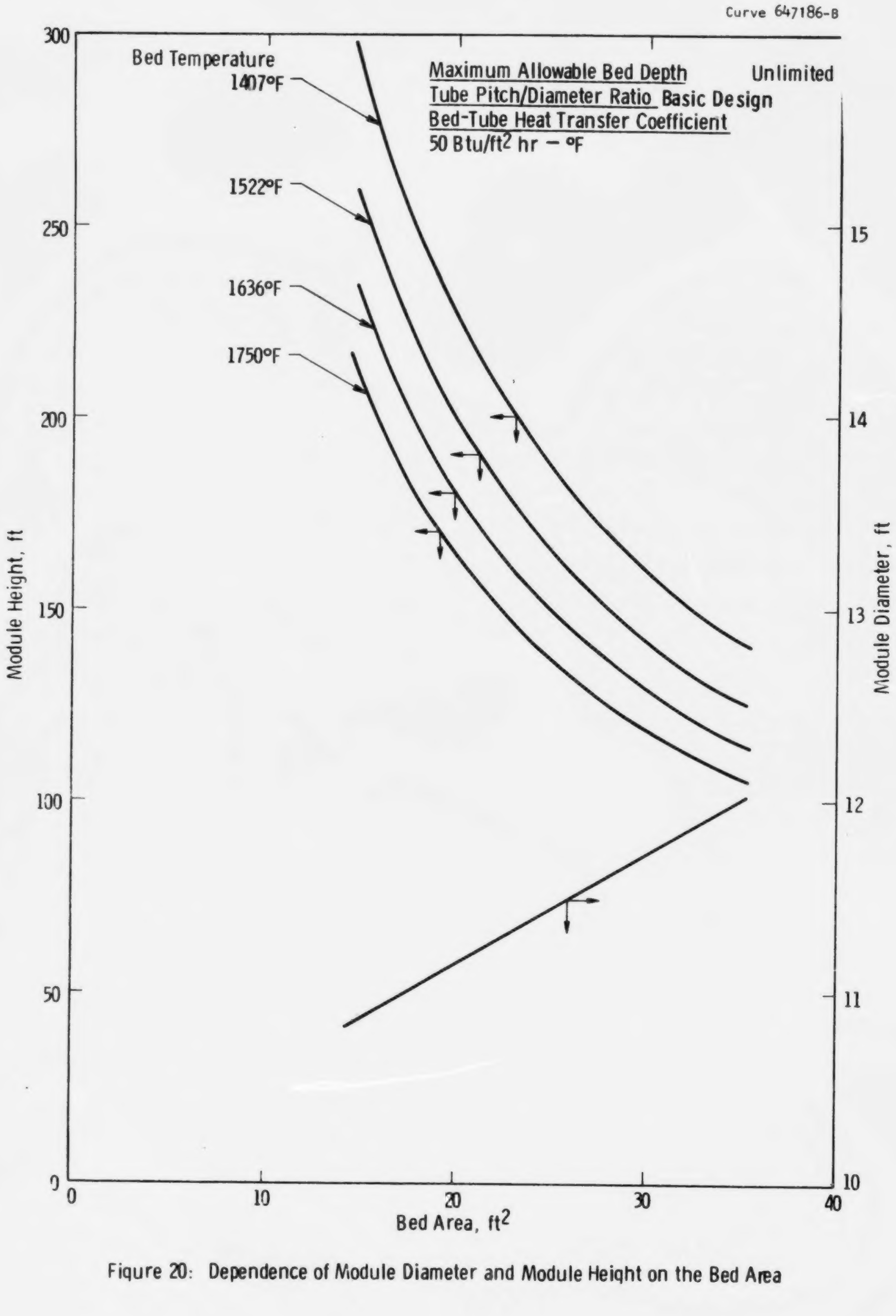


Figure 20: Dependence of Module Diameter and Module Height on the Bed Area

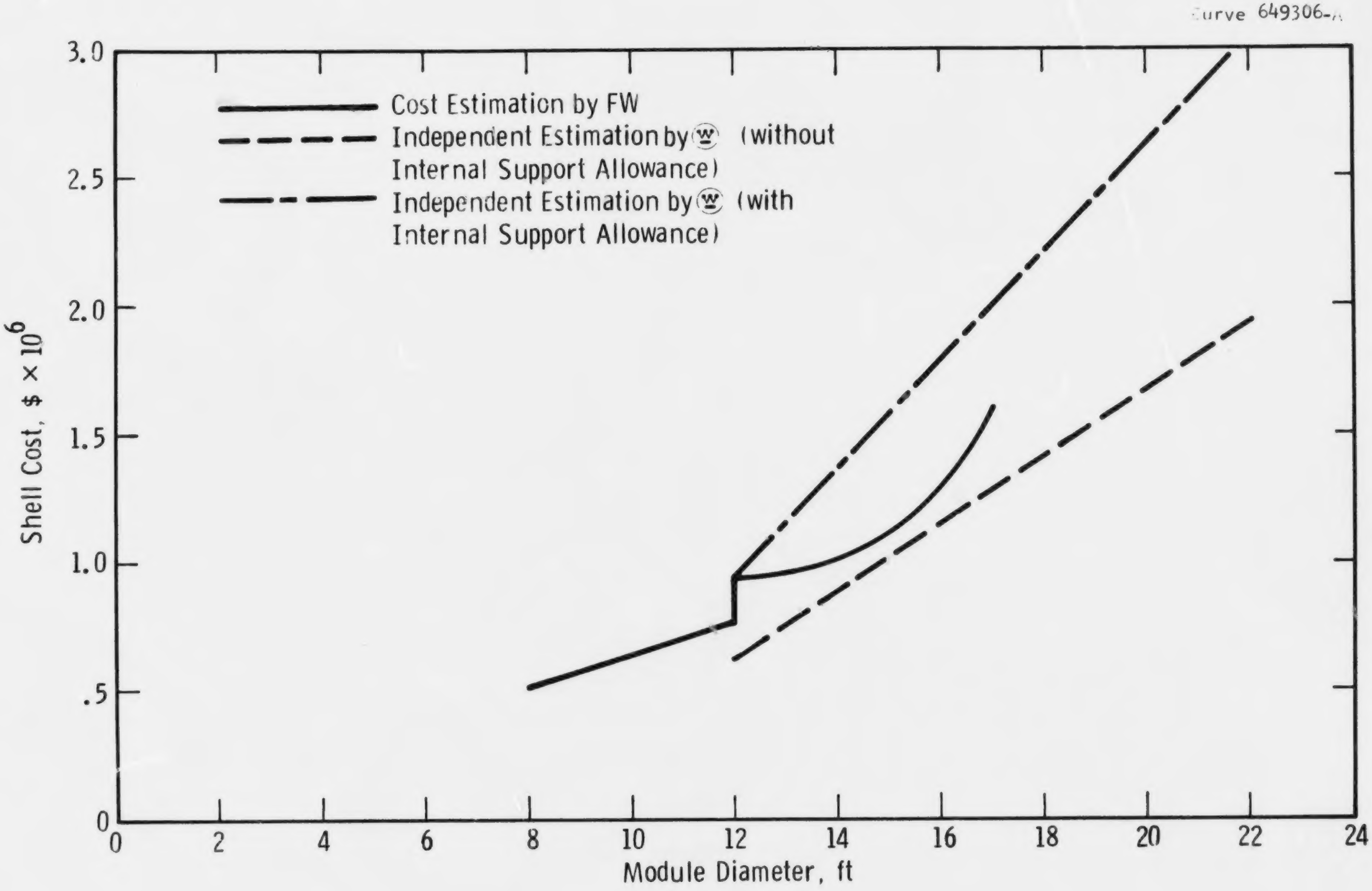


Figure 21: Shell Cost (4 modules) at Basic Design Height

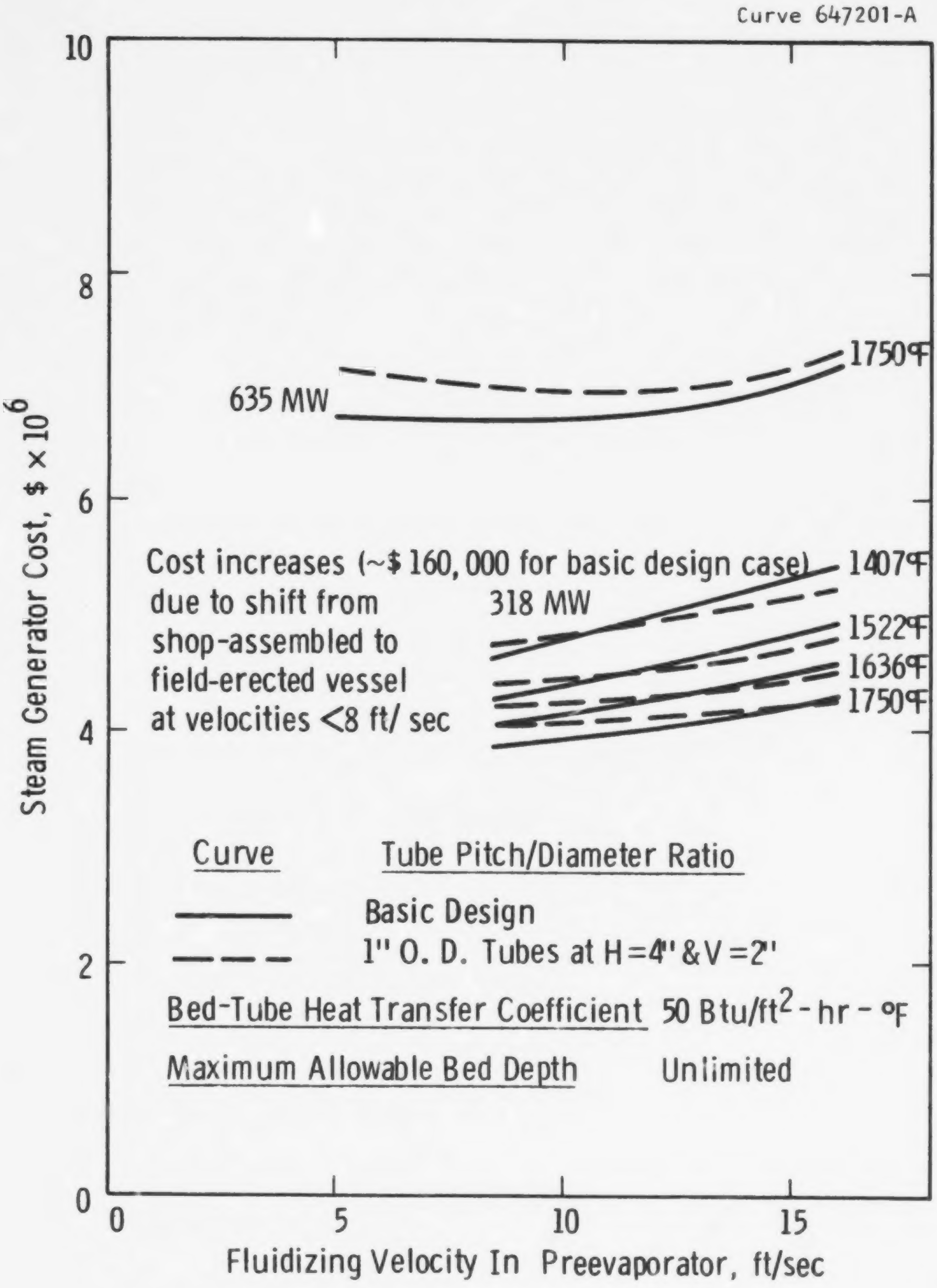


Figure 22: Dependence of the Steam Generator Cost on the Fluidizing Velocity

It is important here to understand why increasing the fluidizing velocity increases the steam generator cost at the 318 MW size and produces a minimum at the 635 MW size. To illustrate the point better, cost reduction due to a decrease in module diameter and cost escalation due to an increase in module height for a four-module design were shown in Figure 23 for the design bed temperature at 1750°F. Increasing the fluidizing velocity escalates the steam generator cost almost linearly from the basic design point due to an increase in module height (curve 2). At the same time, the cost decreases because of a decrease in module diameter; however, the decrease is much more gradual and levels off at a higher fluidizing velocity (curve 1). This is because the bed area alone occupies less than 40% of the total cross-sectional area of a pressurized module. The remaining area is required for piping and headers, and this space is relatively unchanged at a specific plant size even though the bed area is reduced to increase the fluidizing velocity. Thus, the steam generator cost increases with an increase in fluidizing velocity. At the 635 MW size, however, the cost reduction due to a decrease in module diameter is larger during the initial deviation from the basic design point and thus creates a minimum (Figure 23).

Another approach for analyzing the effect of fluidizing velocity would be to change the number of modules as well as the module diameter. Based on turndown consideration, it is preferable to have a four-module-design; however, if a three-module design showed a substantial saving with negligible effect on turndown capability, it would be a better choice. Changing the fluidizing velocity will change the total bed area required for each functional bed, but the total bed volume for each functional bed will remain constant once the tube size and spacing are fixed. Thus, a design with a smaller number of modules would require a larger module diameter at the same design fluidizing velocity. An estimation can usually be performed to evaluate the relative economy between these two designs. For example, consider a four-module and

Curve 647185-A

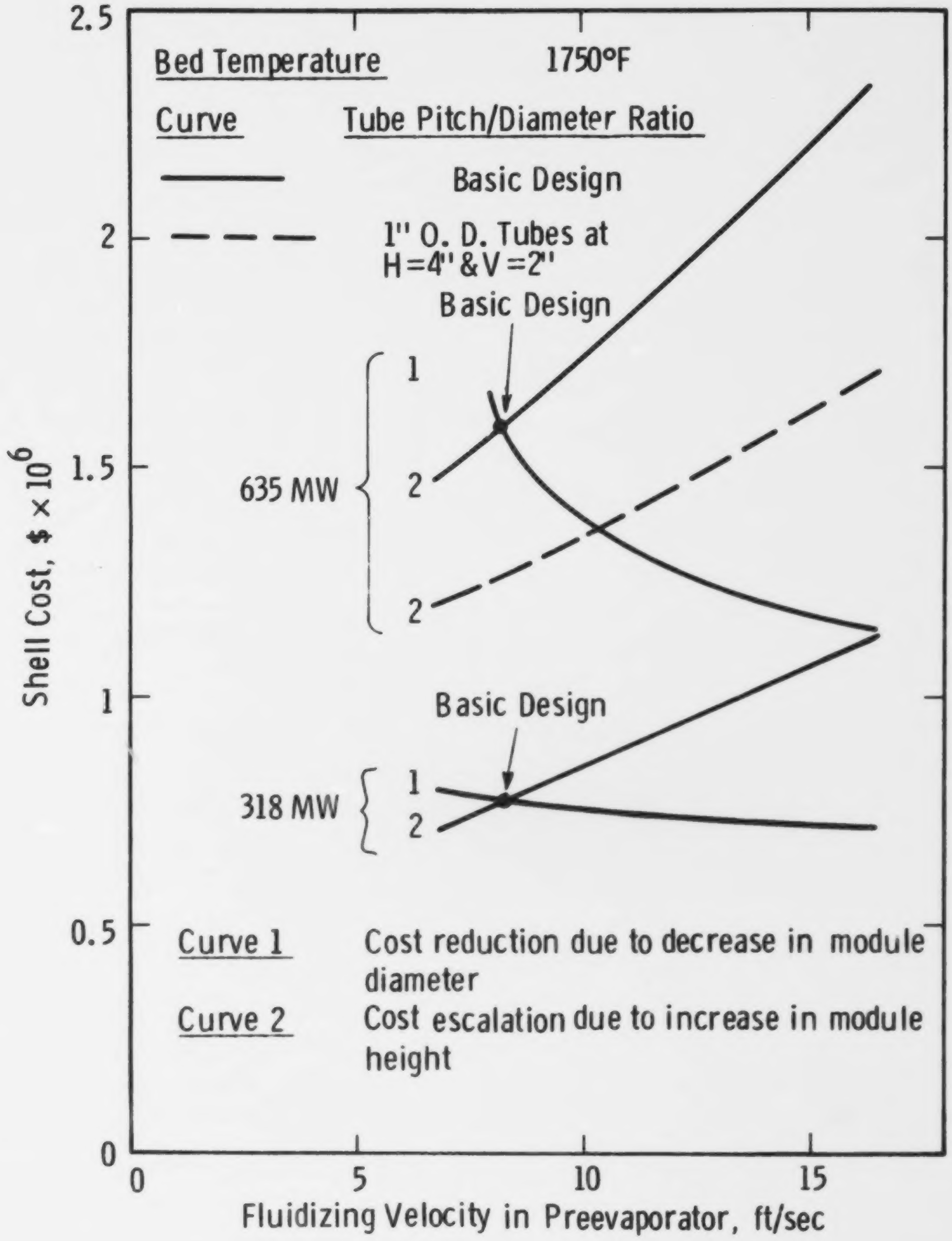


Figure 23: Dependence of the Shell Cost on the Fluidizing Velocity

a three-module design at the same design fluidizing velocity and with the same bed depths. Since the bed volume of each functional bed is constant at fixed tube size and spacing for both cases, we have

$$\frac{D_4}{D_3} = \sqrt{\frac{3}{4}} \quad , \qquad (1)$$

if the bed height is assumed constant for both cases and the bed area occupies a fixed percentage of the total cross-sectional area of the module, where D_4 and D_3 are the respective module diameters for the four-module and three-module designs. Since the shell cost is dependent on the vessel diameter (Figure 21), the relative advantage of these two designs will depend on the plant size in question. For example, for D_4 = 12.5 ft, D_3 can be calculated from Equation 1 to be 14.4 ft. The shell cost can be found from Figure 21 to be \$0.94 x 10^6 for the four-module design and \$0.79 x 10^6 for the three-module design. If the fabrication cost of the module internals is similar in both cases, the three-module design will have a slight economic advantage; however, this advantage becomes progressively smaller because of the rapidly increasing shell cost at large module diameters and the rapidly decreasing portion of the design which could be shop-fabricated. It is estimated that the largest module diameter which is still economical for the three-module design is $\sim$ 17 ft. This conclusion is based on the assumption that boiler turndown is not a problem. If the module diameter is in a range where the module can be shop-fabricated and transported by railroad (i.e., $<$ 12 ft), designing for the maximum shippable module will have definite advantages, provided that turndown is not a problem.

It is concluded from this analysis that for a plant size around 300 MW, the module diameter should be the largest within shipping limitations (12 ft for railroad transportation); at a 600 MW plant size, an optimum fluidizing velocity exists and it should be found for each capacity. However, the cost deviation from that of the optimum design is less than \$1.00/kw (Figure 22). Of course, decreasing the bed area may reduce the number of feed points, but the saving is only ~ \$0.10/kw. Although the effect of fluidizing velocity on the steam generator cost is presented primarily on the basis of the basic design conditions, i.e., the bed-tube heat transfer coefficient = 50 Btu/ft^2-hr-°F, the trends would be the same for a bed-tube heat transfer coefficient at 35 or 75 Btu/ft^2-hr-°F. Change in tube size and spacing may change the slope of the curves or alter the minimum in Figure 17, but the conclusions will remain the same.

The effect of changing the fluidizing velocity on the bed-tube heat transfer coefficient, combustion efficiency, and total particulate carry-over is not taken into account in this analysis because of a lack of accurate quantitative data. The effect of dust loading and particle size distribution on the cost of the particulate removal system is evaluated separately.

2.2.3 Excess Air

A change in design excess air will affect the cycle efficiency and the cost of the boiler module, the steam- and gas-turbine equipment, and the particulate removal system. In order to quantify the effect of excess air on the total boiler cost, the air/fuel ratio is allowed to vary but the total fuel input is kept constant. To simplify the analysis, other parameters -- bed temperature, tube size and tube spacing, bed-tube heat transfer coefficient, fluidizing velocity, and number of boiler modules -- are held constant at the basic design values.

Thus, when excess air is increased beyond the design value, the bed area has to be increased if the fluidizing velocity is kept constant. When the bed area is increased, the module diameter has to be increased as well; however, the module height is decreased because of a decrease in bed depth (Figure 24). Increasing air input to the bed will also increase the amount of heat carried out from the bed by air and reduce the total heat transferred in the bed. At a constant bed-tube heat transfer coefficient, the total heat transfer surface requirement is reduced. At

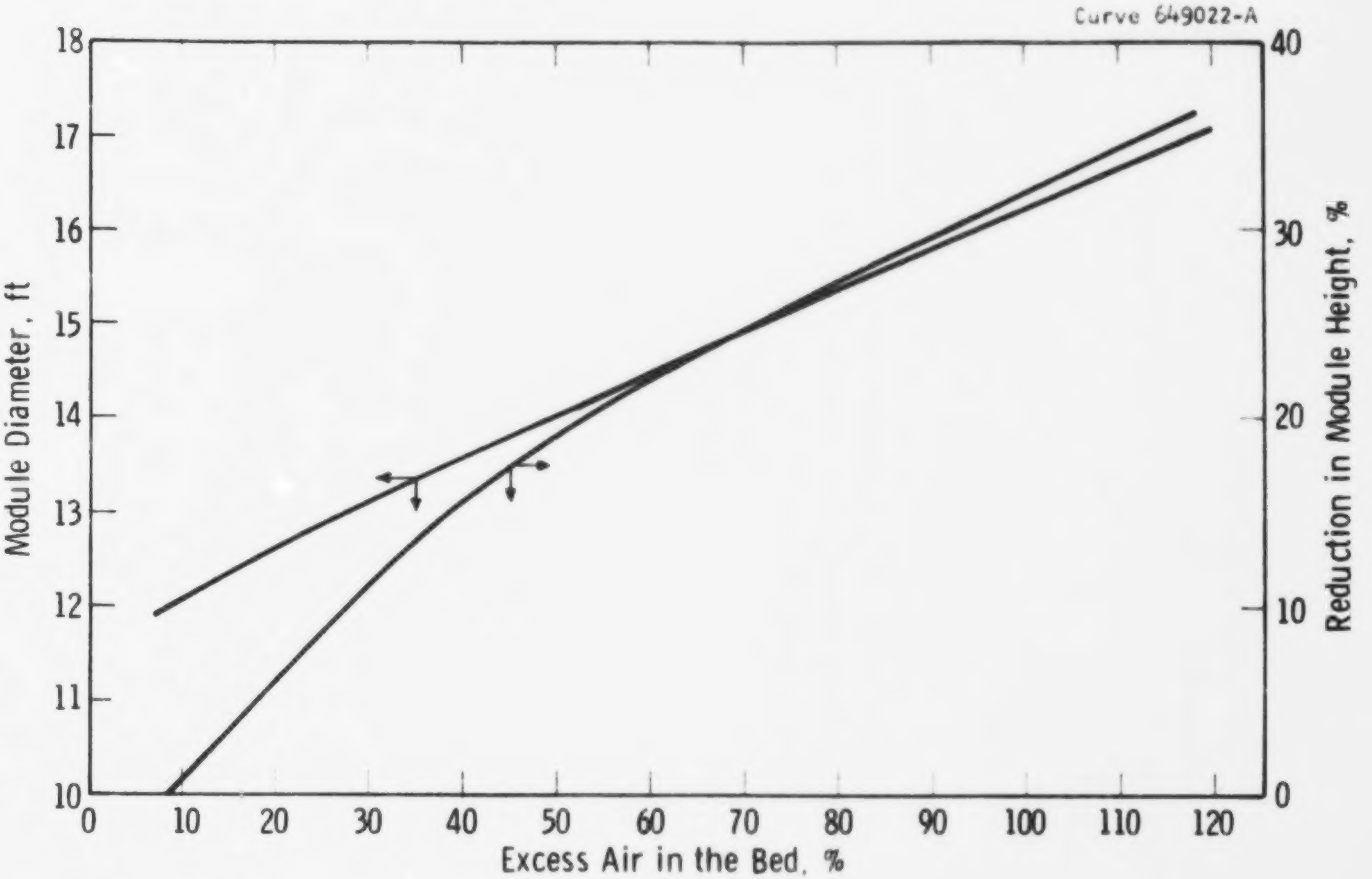


Figure 24: Increase in Module Diameter and Decrease in Module Height at Different Airflow Rate

constant tube size and tube spacing, the bed depth is also reduced, as is the total module height. The bed depth requirement at different airflow rates is presented in Figure 25. At 100% excess air, a reduction of > 40% in the heat transfer surface and a reduction of > 30% in the module height is possible. The bed depths for different functional beds are reduced to ~ 4 ft, which will decrease the pressure drop through the beds and increase cycle efficiency. The module diameter, in turn, is increased from the original 12-ft inside diameter to more than 16 ft for an 80 MW module. Transferring all these changes into economics, an increase in excess air can reduce the boiler cost up to ~ $0.60/kw, as shown in Figure 26. Cost reduction due to the heat transfer surface increases continuously with respect to excess air because of the decrease in the total amount of heat transferred in the bed. Cost reduction due to the pressure shell increases first because of a reduction in module height and then decreases because of the increase in module diameter.

Increasing the excess air will increase the overall plant efficiency as shown in Table 9. Larger gas turbines or additional gas turbines are needed to handle the increased mass flow of gas. If additional units are used, the increase in the gas-turbine equipment cost is shown as a cost adder (Figure 27). This does not account for cost reductions which can be realized by going to larger turbine capacities. This cost increase in gas-turbine equipment is partially offset by a decrease in the steam-turbine equipment cost also shown in Figure 27. The major equipment items taken into consideration in this analysis include gas-turbines with external manifolds, a steam-turbine system, circulating water and condensing systems, a feedwater system including station piping, and stack gas coolers.

Instead of keeping the fluidizing velocity constant, the bed area and module diameter can be kept constant, allowing the fluidizing velocity to increase with excess air. In this case, the cost reduction in the heat transfer surface will be similar, but the cost reduction in the

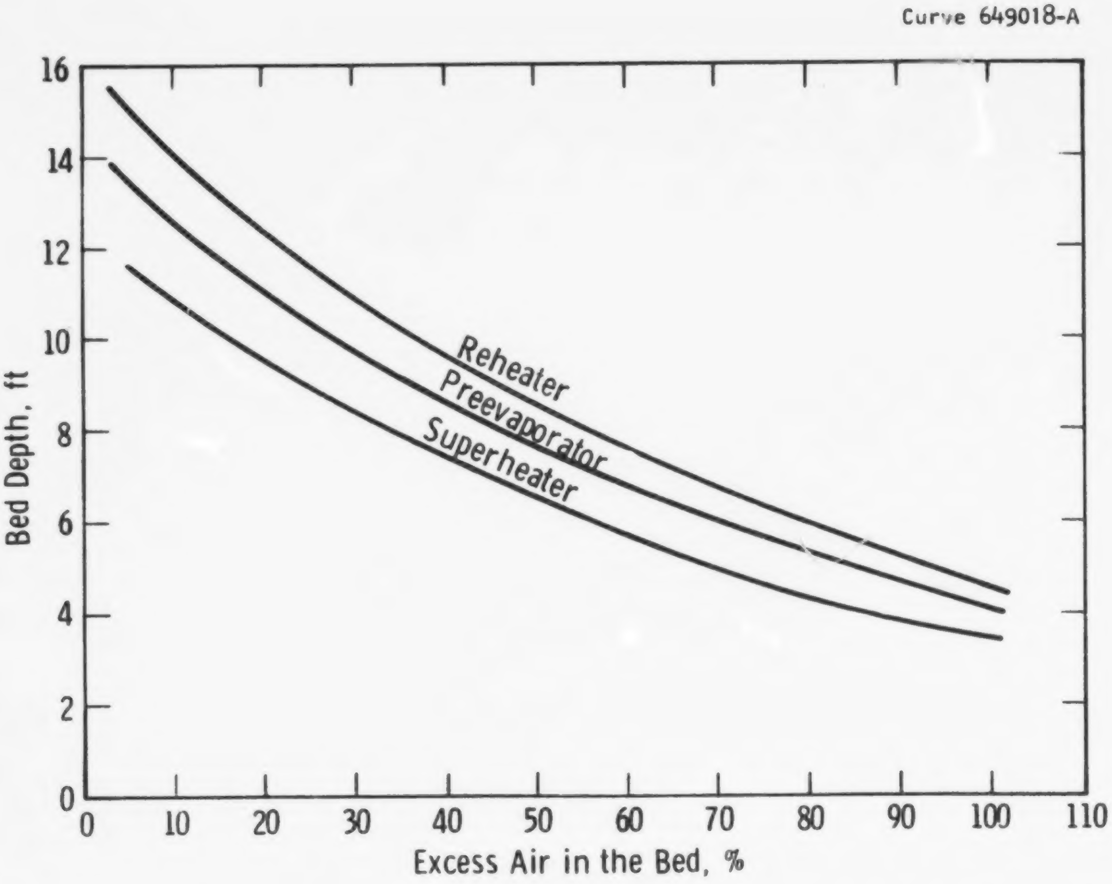


Figure 25: Bed Depth Requirement at Different Airflow Rate

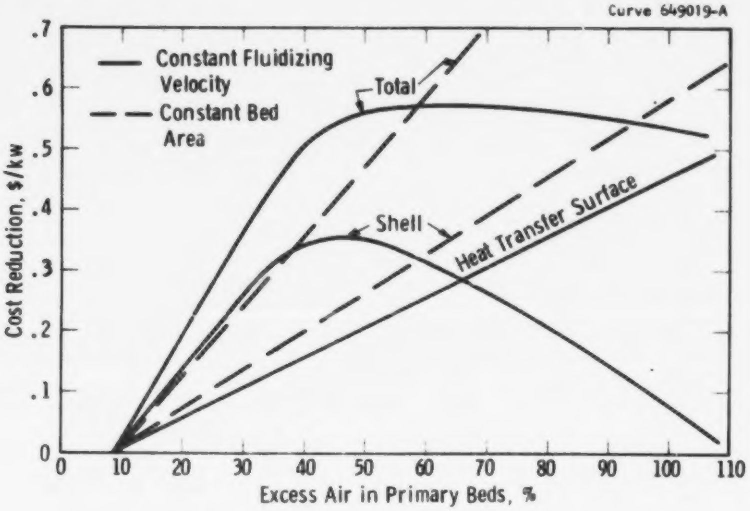


Figure 26: Cost Reduction in Heat Transfer Surface and Pressure Shell at Different Airflow Rates

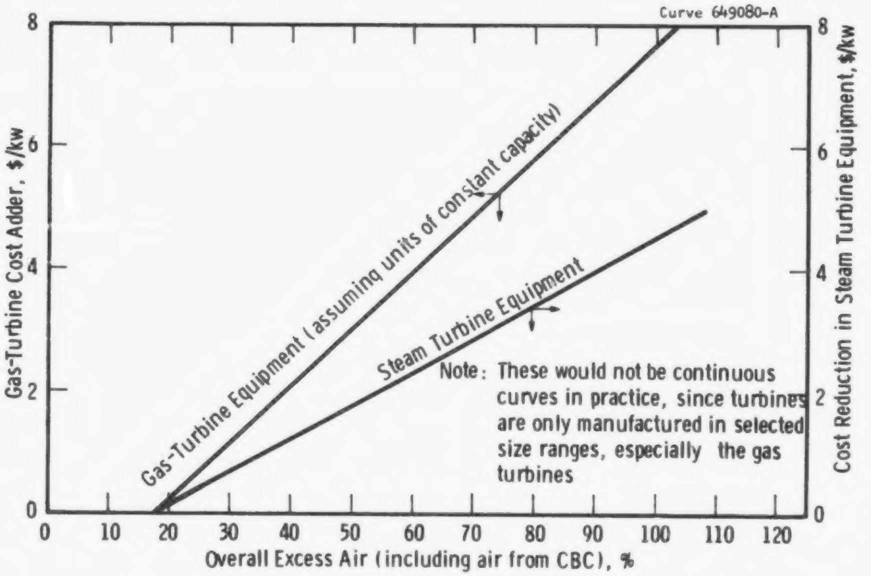


Figure 27: Cost Escalation in Gas-Turbine Equipment and Cost Reduction in Steam-Turbine Equipment at Different Airflow Rates

TABLE 9

EFFICIENCY CALCULATIONS AT VARIABLE AIRFLOW RATES

Excess Air[a] %	F/A Ratio	Plant MW	Heat Rate btu/kWh	Gas Turbine MW	Steam MW	Ratio GT/Steam	No. of Gas Turbine
17.5	0.0861	644.1	9026	127.9	532.2	0.240	2.000
50.0	0.0674	648.0	8972	156.9	506.3	0.309	2.450
90.0	0.0532	649.7	8948	193.5	470.4	0.411	3.025

[a] Excess air is overall excess air, including air from carbon burn-up cell.

shell will continuously increase with increasing excess air and not go through a maximum (Figure 26). The cost reduction in heat transfer surface and pressure shell at 100% excess air in this case (with 10 to 15 fps fluidizing velocity) is estimated to be ∿ \$1.00/kw. However, increasing the fluidizing velocity to larger than 15 ft/sec may be impractical in this design approach.

Increasing the excess air will decrease the total heat transfer surface required in the fluid bed until no boiler tube surface will be required at an excess air of approximately 300%. In this case the power system would become a combined cycle plant, with the gas to the turbine expanders supplied from a coal-fired, adiabatic combustor. The heat recovery boiler would probably be unfired. This system concept differs significantly from a pressurized fluidized bed boiler power plant concept. For example, the boiler becomes an adiabatic combustor, particulate removal equipment costs increase significantly because of the increased gas flow, gas piping costs increase, and the gas-turbine power contribution increases from ∿ 20% up to ∿ 70%. An economic analysis and assessment of this system has been made in Section 6. The heat rate for the adiabatic combustor plant is projected to be 100 to 500 Btu/kWh (depending on the gas-turbine inlet temperature) greater than for the pressurized boiler plant.

Increasing the excess air will provide more flexibility in turndown. At 100% excess air, an additional ∿ 10% load reduction is possible, as compared with operation at 10% excess air. This means a boiler designed at 1600°F and 100% excess air will be able to meet a 4:1 turndown requirement. An adiabatic combustor system should extend the turndown capabilities.

More discussion on excess air- and gas-flow rate appears in the section on particulate removal.

2.2.4 Operating Pressure

The full load design pressure is 10 atmospheres. When the design pressure level is changed with the other operating parameters constant, the gas density will change in proportion to the pressure, and the volumetric flow will vary accordingly. Therefore, the bed cross-sectional area will have to be changed to maintain constant fluidizing velocity, and the bed depth will have to be changed to maintain constant bed volume. The heat transfer coefficient may change because of changes in the quality of fluidization.

The changes in volumetric flow and in gas density will affect the design of the particulate removal equipment. Gas-turbine cycle efficiency is also dependent on the operating pressure. However, analysis of the high-pressure fluidized bed boiler system indicates a reverse direction in pressure level effect. Auxiliary equipment such as the coal and dolomite feeding systems are pressure dependent as well. The relative importance of these factors with respect to operating pressure is analyzed in the following paragraphs.

First consider the boiler module alone. At constant fuel feed rate and excess air, increasing the operating pressure will decrease the gas volumetric flow rate. There are two approaches in designing the boiler modules: keep the bed area constant and let the fluidizing velocity change with the operating pressure; or keep the fluidizing velocity constant and change the bed area according to the operating pressure. The incremental module cost for the constant bed area case is the cost of reinforcing the pressure shells, since for a constant overall heat transfer coefficient and a specific design bed temperature the total heat transfer surface in the bed is constant. This amounts to \$0.20/kw and \$0.30/kw for operating pressures of 15 atmospheres and 20 atmospheres respectively at 300 MW nominal plant size. At 600 MW plant size, the respective cost increments are \$0.10/kw and \$0.20/kw. The credit of decreasing particulate carry-over by operating at lower fluidizing velocities is not taken into account. If the fluidizing velocity is maintained constant, an increase in operating pressure will require a decrease in bed area. Since the total heat transfer surface is constant, a decrease in bed area will require an increase in bed depth at constant tube size and tube spacing and thus an increase in the module height. In this case, the effect of the operating pressure on boiler design and boiler cost while keeping the fluidizing velocity constant is the same as the effect of changing fluidizing velocity with the operating pressure constant. Thus, Figures 16 to 19, 22 and 23 in the section on fluidizing velocity can also be applied for this approach if the coordinate for the fluidizing velocity is changed to (fluidizing velocity at basic design) x (new operating pressure/basic design pressure). For a design bed temperature of 1750°F, the cost increments are $0.80/kw and $1.90/kw for operating pressures of 15 atmospheres and 20 atmospheres respectively at 300 MW plant size. At 600 MW plant size, the cost increments are $0.20/kw and $1.20/kw respectively.

Comparing these two design approaches, the constant area case is the less costly one. Moreover, if basic design bed area and fluidizing velocity are maintained, increasing operating pressure will mean a higher capacity shop-fabricated module (i.e., module diameter $<$ 12 ft). At 15 atmospheres operating pressure, a module of $\sim$ 120 MW capacity can be shop fabricated; at 20 atmospheres pressure, the maximum size module that can be shop fabricated is $\sim$ 160 MW. However, the module height will be considerably increased because of the increase in total heat transfer surface required. The total increase in module height will depend on the heat transfer surface arrangement in the bed.

Increasing the operating pressure will reduce the size of the particulate removal equipment because of the decrease in volumetric flow rate and will reduce the particulate removal efficiency as well because of changes in gas density and viscosity. The increase in operating pressure will also require reinforcement of the containment vessel and the piping and ducting. The savings in the particulate removal equipment by operating at 15 atmospheres and 20 atmospheres are estimated to be $4.0/kw and $6.0/kw respectively for a 300 MW plant size. The major saving comes from the secondary collectors, where maximum single unit capacity is assumed restricted to 30,000 ACFM. An increase in operating pressure will result in fewer units. The cost reduction from operating at higher pressure will be even greater for larger plant sizes or at higher design excess air.

Cycle optimization calculation was performed to evaluate the effect of operating pressure. The parameters studied are: intercooled and nonintercooled compression; gas-turbine compressor pressure ratios from 10 to 30; cycle gas side pressure drops of 3 to 8%. The results are summarized in Figure 23 which gives the plant heat rate for the intercooled and nonintercooled cases. For the nonintercooled case, the best efficiency is obtained at a pressure ratio of 10; and for the intercooled case, the optimum pressure ratio is 15 but with a higher heat rate and a more complex gas turbine. Thus, an increase in operating pressure higher than 10 atmospheres decreases the overall plant efficiency. This decrease, however, is small: ∿ 0.2% at 15 atmospheres.

Weighing the above discussions, the only distinct advantage for operating at pressures higher than 10 atmospheres is the capability of shop-fabricating a large capacity plant, especially at higher design excess air.

2.3 BOILER DESIGN

2.3.1 Heat Transfer Surface Configuration

In the 300 MW design, the heat transfer surface is provided by serpentine tubes where horizontal sections are spaced as shown in Figure 29. Tubes of 2 in. O.D. are used at water walls and are spaced 3-1/2 in. apart. Tubes for preevaporator and superheaters are 1-1/2 in. O.D. and tubes for the reheaters are 2 in. O.D. The tubes can usually be arranged in staggered or rotated diamond arrays; or they can be arranged in a square or rectangular pitch (Figure 30).

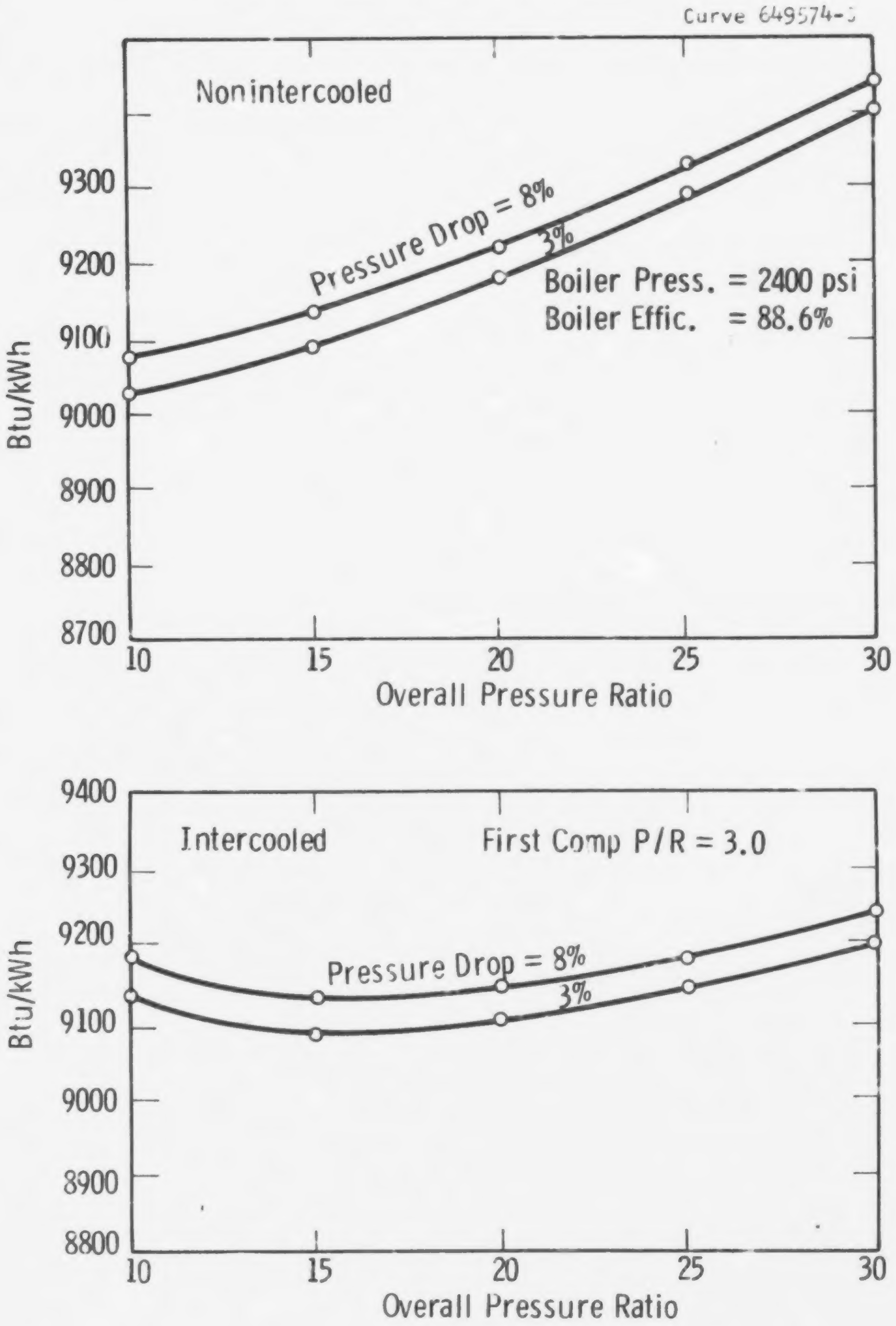


Figure 28: Plant Heat Rate versus Compressor Pressure Ratio

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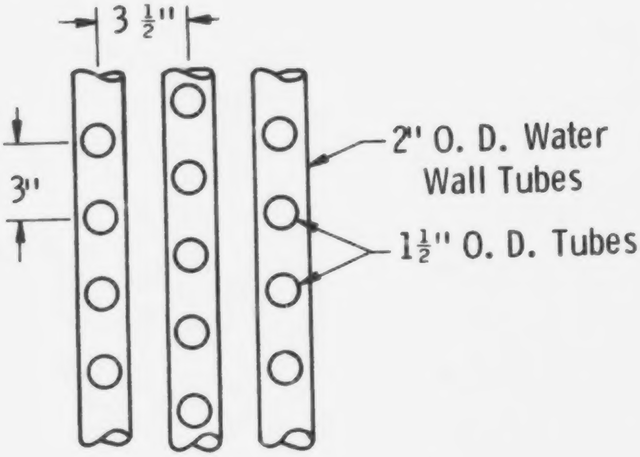
Preevaporator & Superheaters


3 $\frac{1}{2}$"
3"
2" O. D. Water Wall Tubes
$1\frac{1}{2}$" O. D. Tubes

Reheater

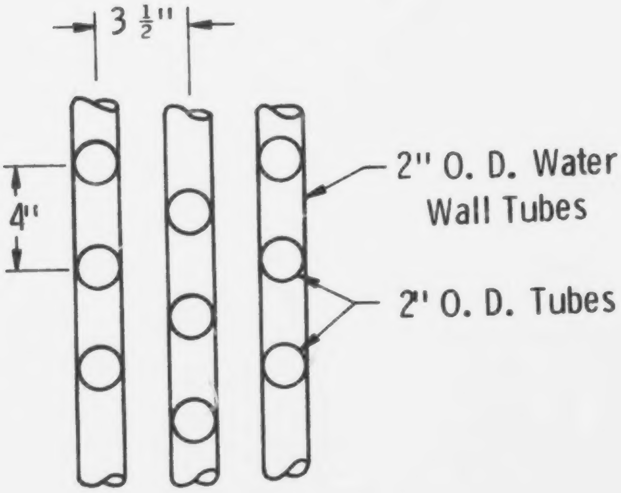

3 $\frac{1}{2}$"
4"
2" O. D. Water Wall Tubes
2" O. D. Tubes

Figure 29: Tube Arrangement in Basic Design

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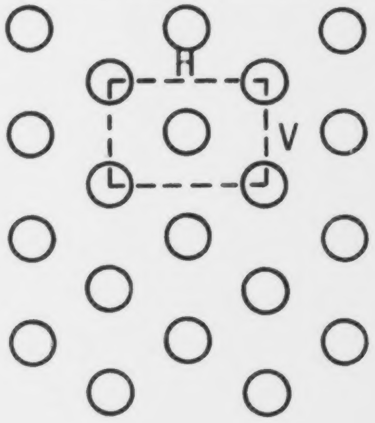
Staggered or Rotated Diamond Arrangement



Square or Rectangular Arrangement

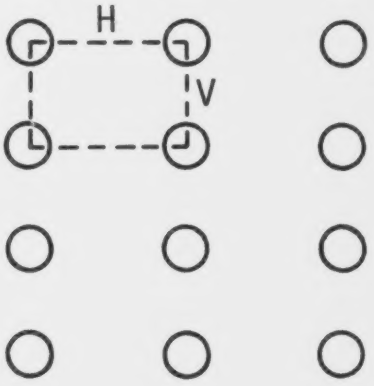


Figure 30: Definition for Different Tube Arrangements

For a staggered or rotated diamond arrangement, the available heat transfer surface per unit bed volume can be expressed as

$$\frac{2\pi D}{HV} = \frac{2\pi}{(\frac{H}{D})\ (\frac{V}{D})\ D} \quad , \tag{2}$$

where

H = horizontal spacing as shown in Figure 30, inches
V = vertical spacing as shown in Figure 30, inches
D = outside diameter of tubes, inches
H/D = horizontal tube pitch/diameter ratio
V/D = vertical tube pitch/diameter ratio .

Keeping the tube diameter (D) constant, the available heat transfer surface per unit volume is inversely proportional to (H/D) · (V/D). For example, decreasing the horizontal and vertical tube pitch/diameter ratio by one-half increases the available heat transfer surface per unit bed volume four times. Keeping the tube pitch/diameter ratios (H/D and V/D) constant, the available heat transfer surface per unit volume is directly proportional to the tube diameter (D). Decreasing the diameter of the tube by one-half, for example, doubles the available heat transfer surface per unit bed volume.

For a square or rectangular arrangement, the available heat transfer surface per unit volume can be expressed as

$$\frac{\pi D}{HV} = \frac{\pi}{(\frac{H}{D})\ (\frac{V}{D})\ D} \quad . \tag{3}$$

With tube diameter (D) and horizontal and vertical tube pitch/diameter ratios (H/D and V/D) constant, the available heat transfer surface per unit bed volume in a square or rectangular arrangement is just one-half that in a staggered or rotated diamond arrangement.

With Equations 2 and 3, the effect of change in tube size and tube pitch/diameter ratio can be easily evaluated. Since the present analysis is based on constant fuel feed rate and excess air, the total heat transfer surface requirement is constant at constant design bed temperature and constant heat transfer coefficient. Change in tube size and tube pitch/diameter ratio will only affect the design in bed depth and module diameter. When tube size and tube spacings are changed from those in the basic design, the bed depth in every functional bed deviates from that in the basic design as well. The amount of deviation for different tube sizes and tube spacings was calculated and expressed as multipliers in Figures 31 through 35. To obtain the new expanded bed depth, simply multiply the bed depth in the basic design by the appropriate multiplier. However, in estimating the cost, a small correction is sometimes necessary because when the bed depth is increased, more and more heat transfer surface at the water walls is immersed in the bed. To keep the bed temperature constant, the heat transfer surface in the bed has to be reduced proportionally to keep the total immersed heat transfer surface constant.

The bed depth correction is estimated as follows: the heat transfer surface at the water wall enclosure per ft bed height can be expressed as

$$(\text{H.T.S.})_W = \left(\frac{12A}{S_W} + \frac{12B}{S_W}\right) \cdot \frac{\pi D_W}{12} \quad . \tag{4}$$

The heat transfer surface as immersed tubes per ft bed height is

$$(\text{H.T.S.})_B = \left(\frac{24A}{H} \cdot \frac{12}{V}\right) \cdot B \cdot \frac{\pi D_t}{12} \tag{5}$$

where

A,B = bed dimensions, ft
S_W = water wall tube spacing, in.
D_W = water wall tube diameter, in.
D_t = immersed tube diameter, in.
H = horizontal tube spacing as defined in Figure 30
V = vertical tube spacing as defined in Figure 30.

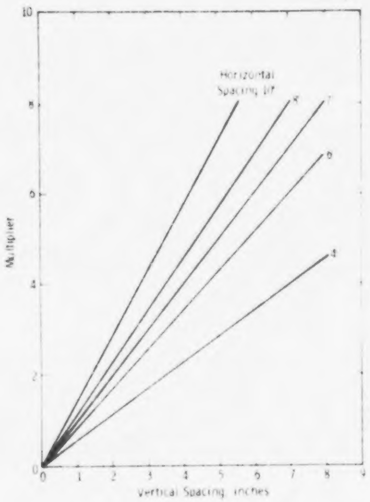


Figure 31 Bed Depth Multiplier for Preevaporator, Superheater, and Reheater (Tube Diameter 1/2" O.D.)

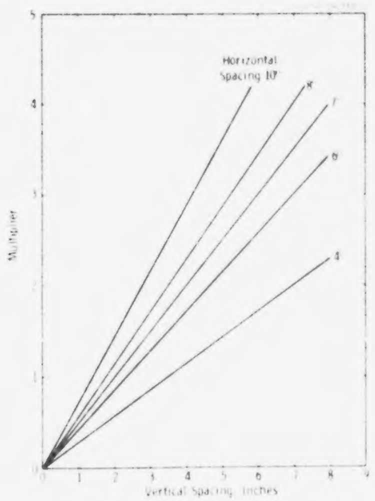


Figure 32 Bed Depth Multiplier for Preevaporator, Superheater, and Reheater (Tube Diameter 1" O.D.)

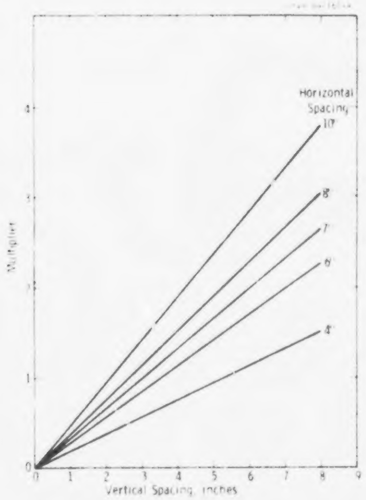


Figure 33 Bed Depth Multiplier for Preevaporator, Superheater, and Reheater (Tube Diameter 1-1/2" O.D.)

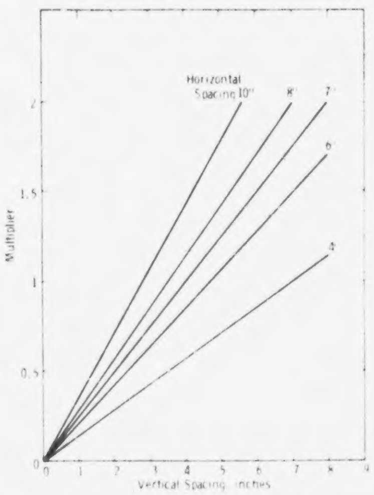


Figure 34 Bed Depth Multiplier for Preevaporator, Superheater, and Reheater (Tube Diameter 2" O.D.)

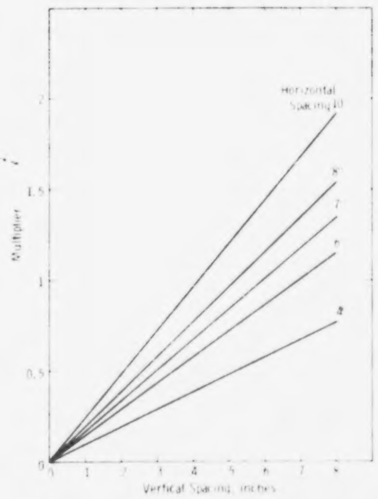


Figure 35 Bed Depth Multiplier for Preevaporator, Superheater, and Reheater (Tube Diameter 3" O.D.)

Dividing Equation 5 by Equation 4 gives the ratio of heat transfer surface as immersed tubes to heat transfer surface of the water wall enclosure per ft bed height, R.

$$R = 24 \cdot \left(\frac{A \cdot B}{A + B}\right) \cdot \left(\frac{S_W}{H \cdot V}\right) \cdot \left(\frac{D_t}{D_W}\right) \qquad (6)$$

Once R is known, the bed height correction can readily be made, but the correction is usually less than 5% of the bed depth in question.

This redistribution of heat transfer surface also requires a redistribution of heat duties because now more heat is transferred to the water walls. Thus water walls, in addition to the evaporation duty, may take up some superheating duty. The amount of redistribution of heat duties depends on the amount of change of bed area from that in the basic design. The smaller the bed area,the smaller the ratio of heat transfer surface in the bed per unit bed height to immersed heat transfer surface at water walls per unit bed height and, thus, the higher the amount of heat duty redistribution.

The effect of the tube pitch/diameter ratio on the cost of the steam generator (not including erection) for a constant 12-ft module diameter was evaluated for three different tube sizes and tube spacings with respect to change in the design bed temperatures; the results are plotted in Figure 13. Curve 2 represents the estimated cost for a staggered arrangement of 1-in. O.D. tubes at H = 4 in. and V = 2 in. (see Figure 30 for definition) in all beds. Curves 3a and 3b are for a staggered arrangement of 2-in. O.D. tubes at H = 8 in. and V = 4 in. in all beds. Some interesting trends are present if these results are compared to the steam generator cost for the basic design. Ignore curves 1b and 3b for the time being because the assumption of unrestricted maximum allowable bed depth is not considered reasonable. Then,

- Decreasing tube size and tube spacing increases the steam generator cost at design bed temperatures above ∿ 1520°F (curve 2). A further decrease in bed temperature necessitates splitting the reheater bed in the basic design into two beds and substantially increases the steam generator cost of the basic design.

- Increasing tube size and tube spacing also increases the steam generator cost (curve 3a).

The reasons for these results are as follows:

- When tube size and tube spacing are decreased, more heat transfer surface can be immersed in a unit bed volume, which results in lower bed height and module height; thinner wall tubes can be used, resulting in lower tubing cost. These are positive advantages.
- Smaller tube size and tube spacing, however, increase the amount of tube bending and tube welding required. This is because more tubes of smaller diameter are required to carry the same water/steam load at a constant flow rate in the tube. Fabrication cost as a function of tube wall thickness is not taken into account because not enough information is available. Pumping costs, a part of total operating cost, are not included here.

The balance of these two factors determines the total steam generator cost.

Since the tubing cost constitutes only about 20% of the cost of the pressure parts, the increase in fabrication cost is more important. This can best be illustrated in Figure 36 where the component costs are plotted against the available heat transfer surface per unit volume which relates to the tube pitch/diameter ratio. The shell cost increases almost linearly with decreasing heat transfer surface per unit bed volume. The cost of tubing, headers, downcomers, and risers is almost constant at heat transfer surface larger than 6 ft^2 per ft^3

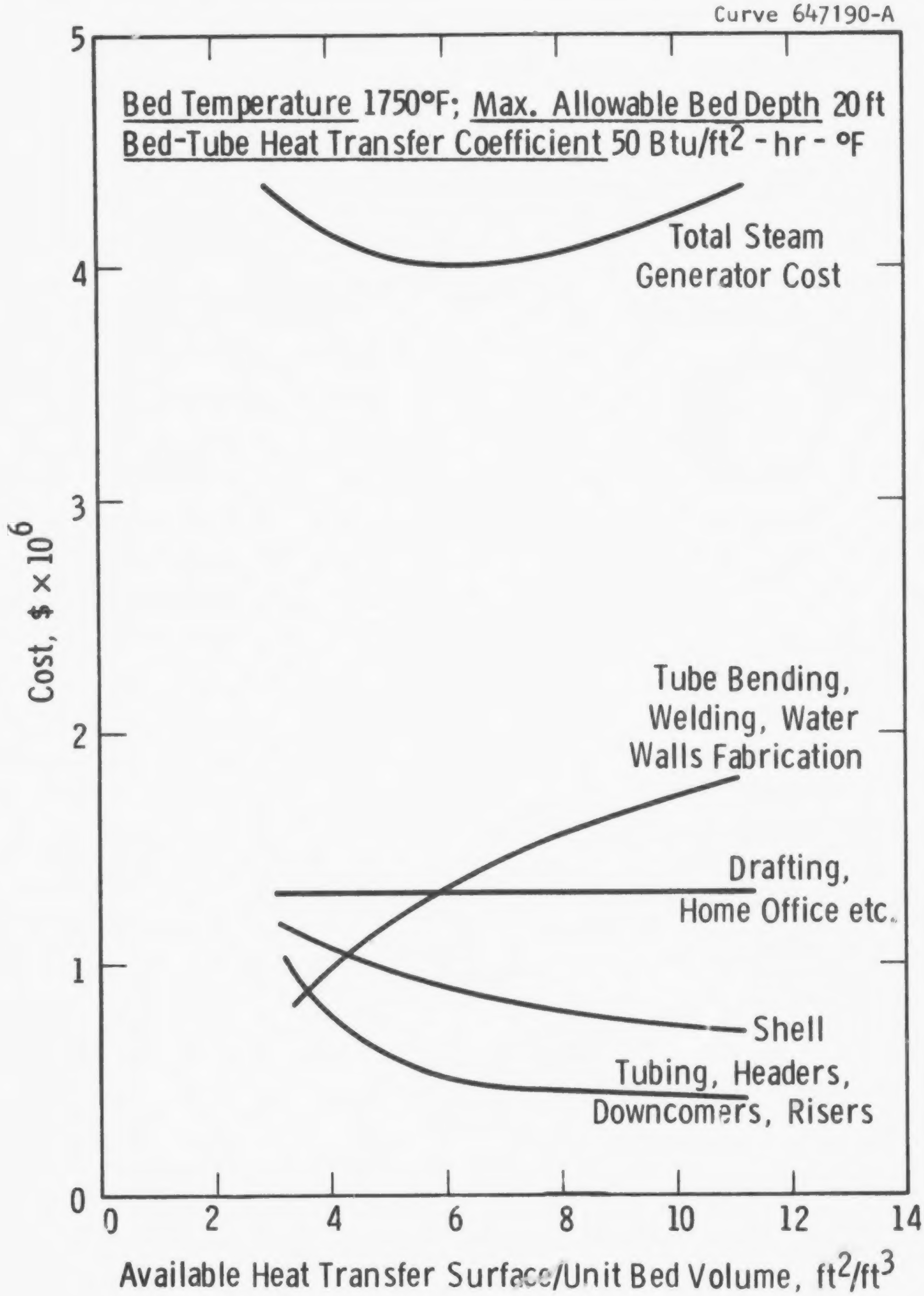


Figure 36: Change of the Steam Generator Cost with Heat Transfer Surface per Unit Bed Volume (not including erection)

bed volume, but increases rapidly at heat transfer surface lower than ~ 6 ft^2/ft^3 bed volume (which corresponds to the use of a tube diameter of 1-1/2" O.D. or larger). When larger tubes are used, the minimum wall thickness increases rapidly and so does the tubing cost which contributes most of the cost escalation at lower heat transfer surface per unit bed volume. However, the cost of the tube bending, tube welding, and water-wall fabrication increases steadily with an increase in heat transfer surface per unit bed volume. The balance of all these factors create a minimum in total steam generator cost at about 6.5 ft^2 heat transfer surface per ft^3 of bed volume. This can be achieved by arranging 1-in.O.D. tubes at H = 4 in.and V = 3 in. Fortunately, the tube size and tube spacing used in the basic design is very close to this actual minimum. Clearly, there are different minimums at different design bed temperatures. An optimum design for a specific operating condition requires a separate evaluation. However, this optimum design point is not as critical as may be generally conceived. For example, at a design bed temperature of 1750°F (Figure 36), the difference in the steam generator cost between the optimum design and other designs is within $1.00/kw for available heat transfer surface of 3 to 11 ft^2/ft^3 bed volume, which covers the three tube sizes and tube spacings in the current evaluation.

The maximum allowable bed depth is a more important variable. The steam generator costs are also plotted against the maximum allowable bed depth at constant design bed temperature and constant bed area in Figures 37 to 40. Although the steam generator of 1-in.O.D. tubes at H = 4 in.and V = 2 in.is not economical when compared to the basic design, it becomes progressively more attractive at a lower maximum allowable bed depth. At maximum bed depths lower than 14 ft (Figure 37), a steam generator using 1-in.O.D. tubes is actually cheaper than the basic design by up to $\sim$ $3.1/kw. At a bed temperature of 1407°F, where more heat transfer surface has to be immersed in the bed, the advantage of using smaller tube size and tube spacing is even clearer (Figure 40). At a

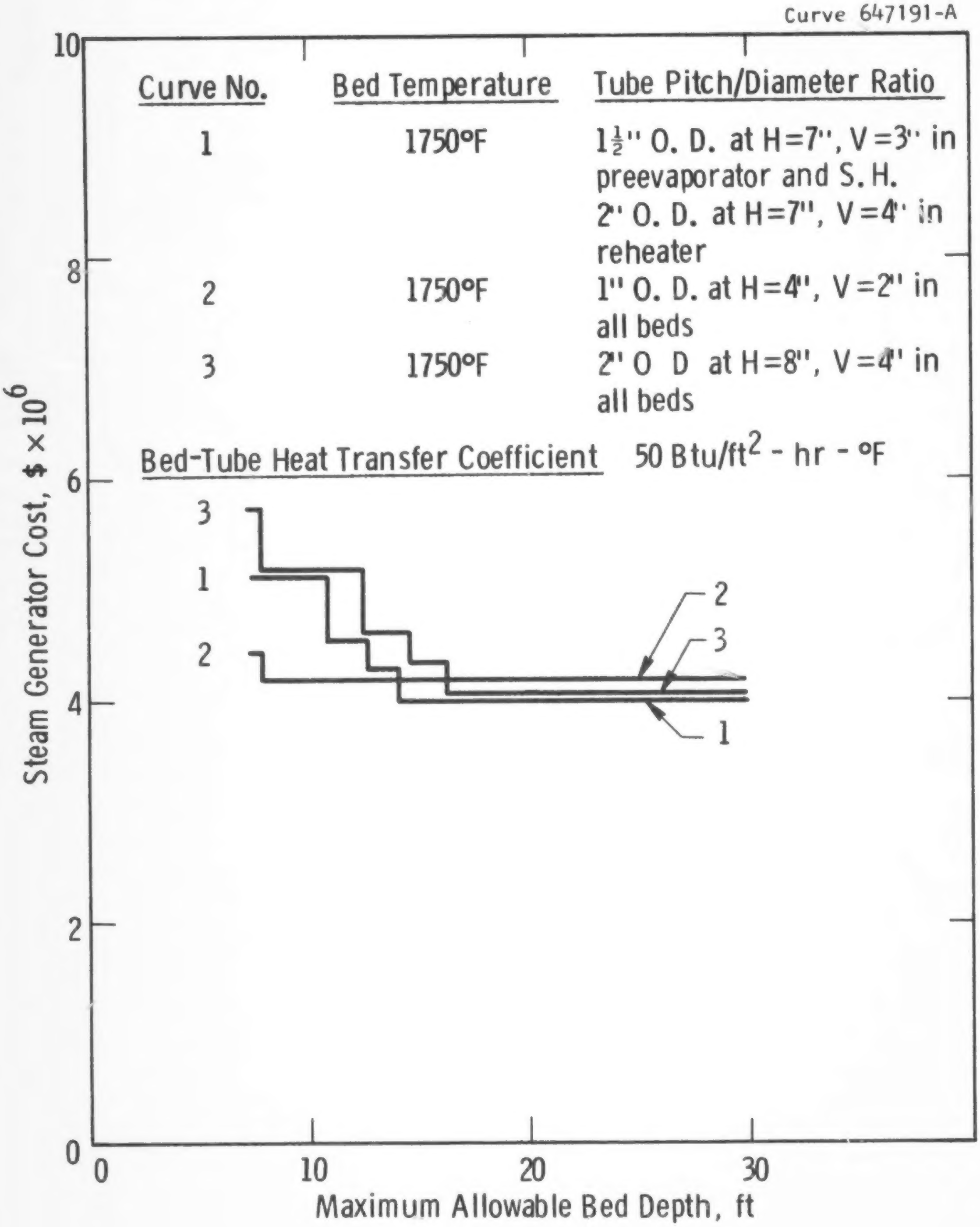


Figure 37: Dependence of the Steam Generator (318 MW) Cost on the Maximum Allowable Bed Depth (not including erection)

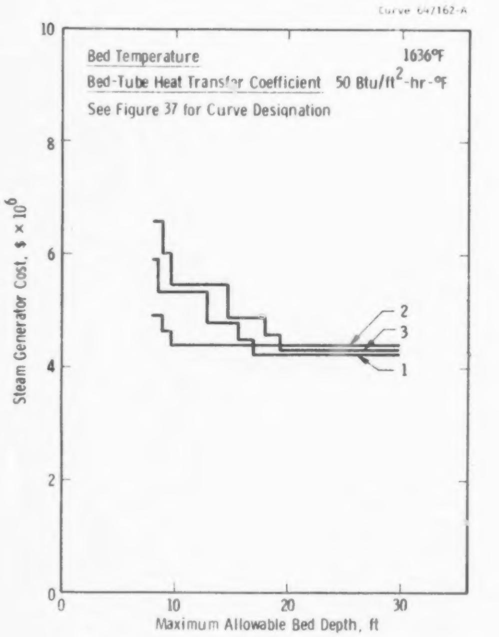


Figure 38: Dependence of the Steam Generator (318 MW) Cost on the Maximum Allowable Bed Depth (not including erection)

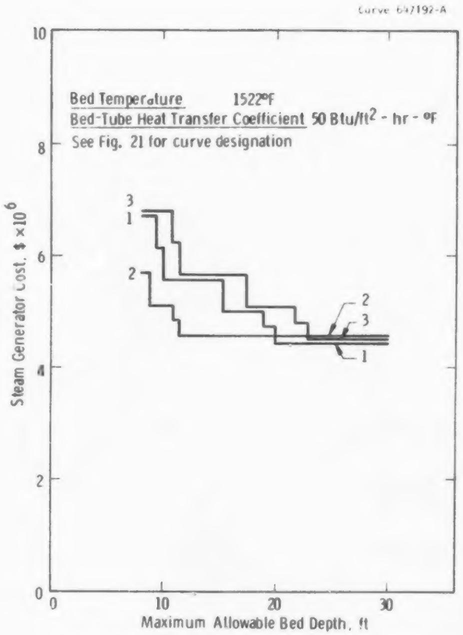


Figure 39: Dependence of the Steam Generator (318 MW) Cost on the Maximum Allowable Bed Depth (not including erection)

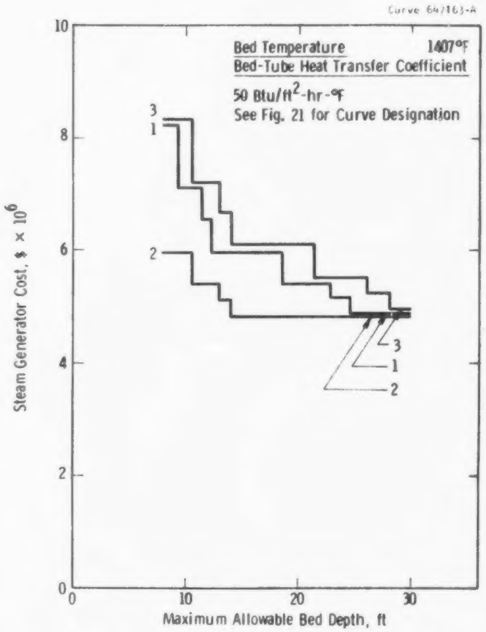


Figure 40: Dependence of the Steam Generator (318 MW) Cost on the Maximum Allowable Bed Depth (not including erection)

maximum allowable bed depth of 10 ft, a $3.1/kw saving over the cost of the basic design is realizable by using 1-in.O.D. tubes at a design bed temperature of 1750°F and a $3.3/kw saving at a design bed temperature of 1636°F.

In this cost evaluation, the credit for using the thinner wall tubes in smaller tubes was also taken into consideration. Minimum wall thickness was calculated for each design bed temperature and was summarized in Tables A-1 and A-2 of Appendix A for both 1-in.O.D. and 2-in. O.D. tubes.

In conclusion, although the use of smaller tube spacings will increase the amount of heat transfer surface immersed in a unit bed volume, the economy is not always favorable because the accompanying increase in cost for tube bending, tube welding, and fabrication sometimes overtakes the savings in tubing cost and module height. However, smaller tubes and tube spacings do show more advantages when the maximum allowable bed depth is limited. Moreover, there is a definite minimum when cost is plotted against the heat transfer surface per unit volume. This can be utilized to find the optimum tube size and tube spacings.

In some cases, the tube pitch/diameter ratio effect on the bed-tube heat transfer coefficient has to be taken into account in evaluating the optimum tube size and tube spacings. Data indicate that smaller tube spacing tends to decrease the heat transfer coefficient[3,4,6] as shown in Figures 41 and 42. Lowering the pitch/diameter ratio from 8 to 2 results in an increase in the heat transfer surface per unit volume of bed by a factor of 16, and decreases the heat transfer coefficient by only 18% (Figure 41). Thus, the major consideration in determining the pitch/diameter ratio is not the heat transfer coefficient, but the compactness and the cost of the boiler design. There is a gap between tubes, however, depending on the particle size, below which a sharp drop of heat transfer coefficient will occur (Figure 42). Determination of this minimum gap requires further experimental studies.

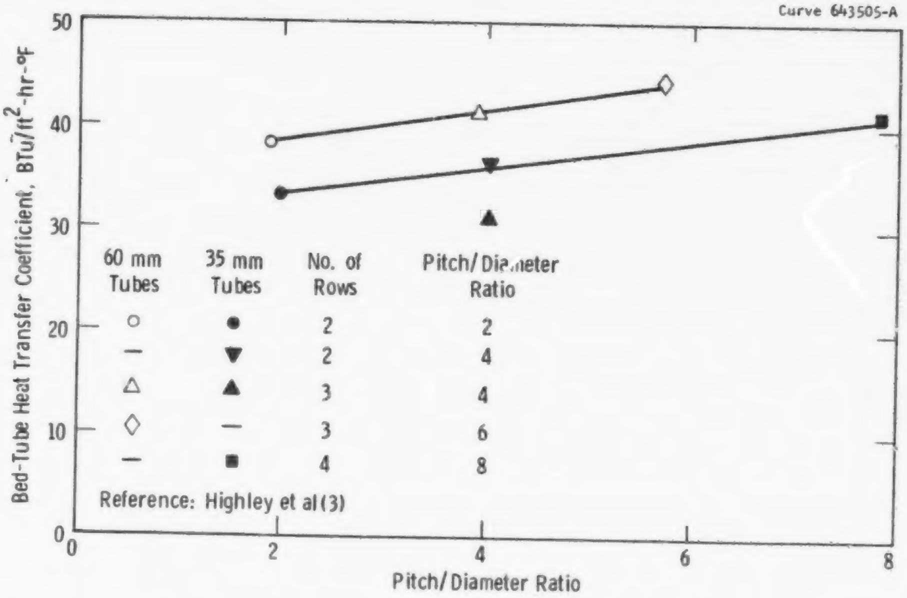


Figure 41: Change of Bed-Tube Heat Transfer Coefficient with Pitch/Diameter Ratio

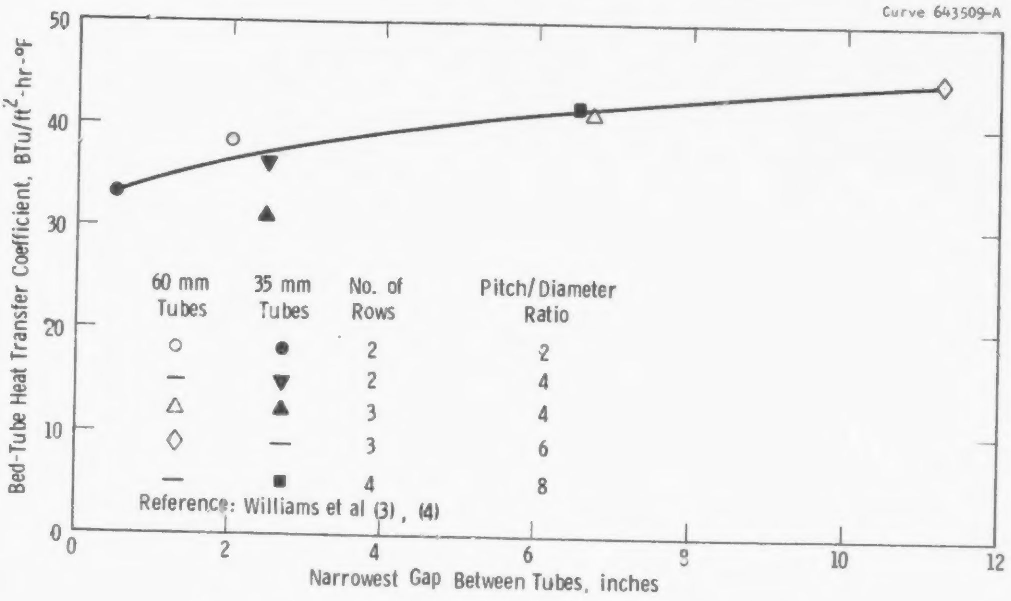


Figure 42: Change of Heat Transfer Coefficient with Tube Spacing

Although this conclusion is drawn from the experimental evidence at atmospheric pressure, it should be qualitatively valid at pressure. Heat transfer coefficients for the tubes at different bed depths may also be different for the pressurized conditions, but unfortunately there is no quantitative information available.

An added advantage to the use of smaller tube sizes and tube spacings is that lower bed depth tends to decrease the pressure drop through the fluid beds and thus increases the gas-turbine efficiency (curve 2, Figure 15).

2.3.2 Heat Transfer Coefficient

In the basic design, the overall heat transfer coefficients assumed are 47 Btu/ft^2-°F-hr for the preevaporator, 45 Btu/ft^2-°F-hr for the superheater, and 43 Btu/ft^2-°F-hr for the reheater. The bed-tube heat transfer coefficient is assumed to be 50 Btu/ft^2-°F-hr for all beds. When the bed-tube heat transfer coefficient is changed, the overall heat transfer coefficient will be changed as well. This leads to a change in the total heat transfer surface requirement and the bed depth. The heat transfer surface and the bed depth multipliers for preevaporator, superheater, and reheater were calculated and are shown in Figure 43. The bed depth thus obtained will have to be adjusted, because when the bed depth increases, the heat transfer surface, as water walls immersed in the bed, increases as well. To keep the bed temperature constant, a corresponding decrease in heat transfer surface in the bed (excluding water walls) is necessary in order to maintain constant total heat transfer surface immersed in the bed. Equation 6 can be used for this correction.

A change in the heat transfer coefficient will also change the tube metal temperature. A change of tube material may be necessary in some cases. The design metal temperature is assumed to be the maximum outside tube wall temperature based on the minimum permissible wall thickness. The minimum wall thickness for bed-tube heat transfer coefficients of 35, 50, and 75 Btu/ft^2-hr-°F is presented in Appendix A.

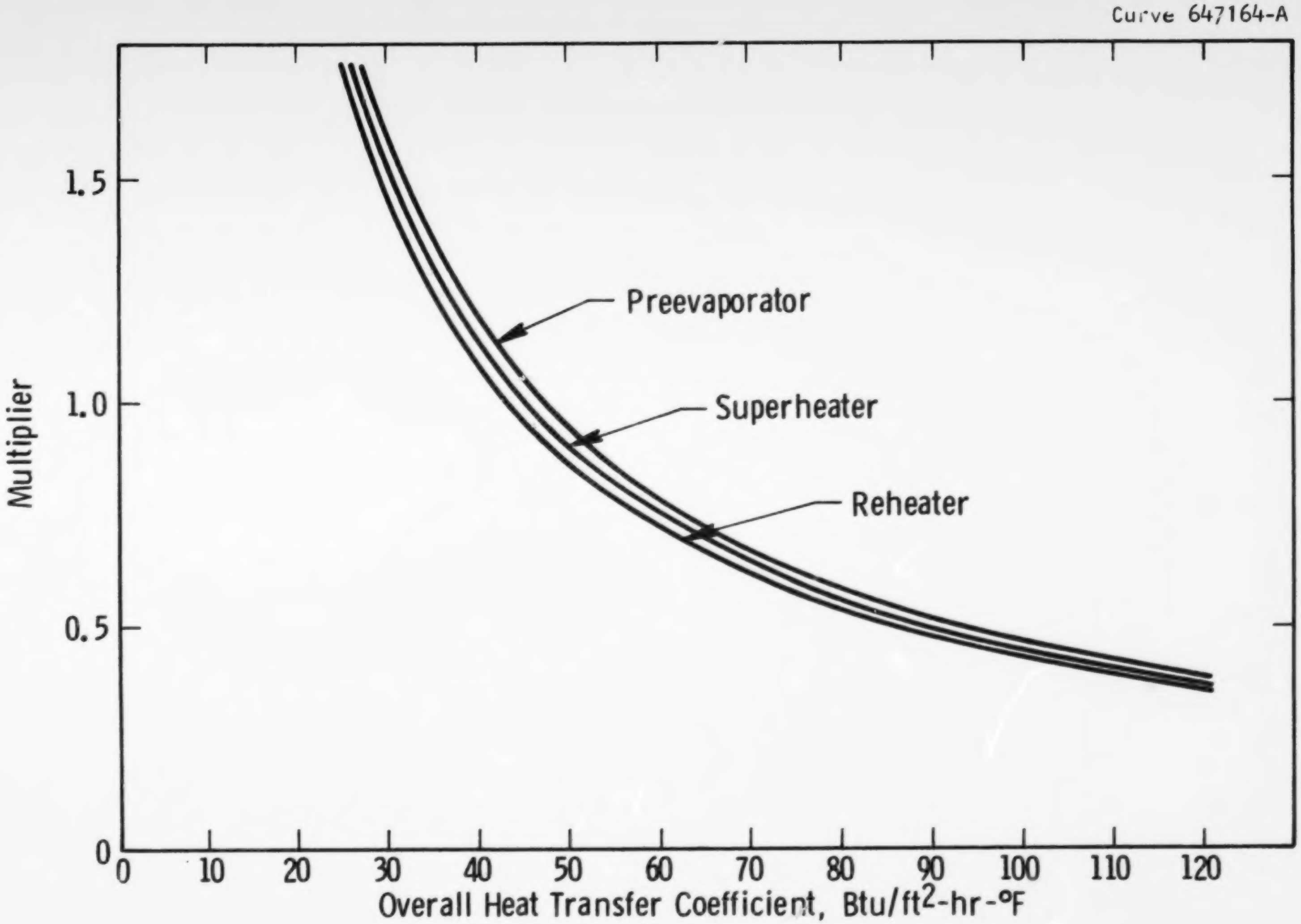


Figure 43: Heat Transfer Surface and Bed Depth Multiplier for Preevaporator, Superheater, and Reheater

Taking into consideration the aforementioned factors, the total steam generator costs at bed-tube heat transfer coefficients of 35 and 75 Btu/ft^2-hr-°F were projected for different tube pitch/diameter ratios at different design bed temperatures (Figures 44 and 45). At a low bed-tube heat transfer coefficient (35 Btu/ft^2-hr-°F) where a large amount of in-bed heat transfer surface is required, a boiler with smaller tube size and tube spacing is much more economical than the one with larger tube size and tube spacing. Savings up to 20% of the total steam generator cost are realizable if the bed area is constant and the maximum allowable bed depth is 20 ft. If the maximum allowable bed depth is less than 20 ft, the saving will be even larger. At a bed-tube heat transfer coefficient of 75 Btu/ft^2-hr-°F, where the heat transfer surface requirement is substantially reduced, smaller boiler tubes and spacings do not have a clear advantage (Figure 45). If the bed-tube heat transfer coefficient is 35 Btu/ft^2-hr-°F, rather than 50 Btu/ft^2-hr-°F as assumed in the basic design, the steam generator cost will increase by \$2.7/kw at 1750°F design bed temperature and by \$5.4/kw at 1600°F. If 1-in. O.D. tubes are used at H = 4 in. and V = 2 in., the cost escalation would be \$2.0/kw at 1750°F, \$3.4/kw at 1600°F, and \$6.1/kw at 1400°F (Figure 44). If the heat transfer coefficient is increased to 75 Btu/ft^2-hr-°F, the reduction in steam generator cost from that of the basic design is only marginal, $\sim$ \$1.0/kw at 1750°F (Figure 45).

Figures 46 and 47 show the effect of the bed-tube heat transfer coefficient on the steam generator cost at constant bed temperature. An increase of the bed-tube heat transfer coefficient from 50 to 75 Btu/ft^2-hr-°F (a 50% increase) decreases the steam generator cost by $\sim$ 10% ($\sim$ \$1.40/kw). A decrease of the bed-tube heat transfer coefficient from 50 to 35 Btu/ft^2-hr-°F (a 30% decrease) increases the cost by $\sim$ 20% ($\sim$ \$2.80/kw) (Figure 46). The curves start to level off at higher bed-tube heat transfer coefficients. Thus, a further increase in the bed-tube heat transfer coefficient larger than about 75 Btu/ft^2-hr-°F does not affect the cost substantially. However,

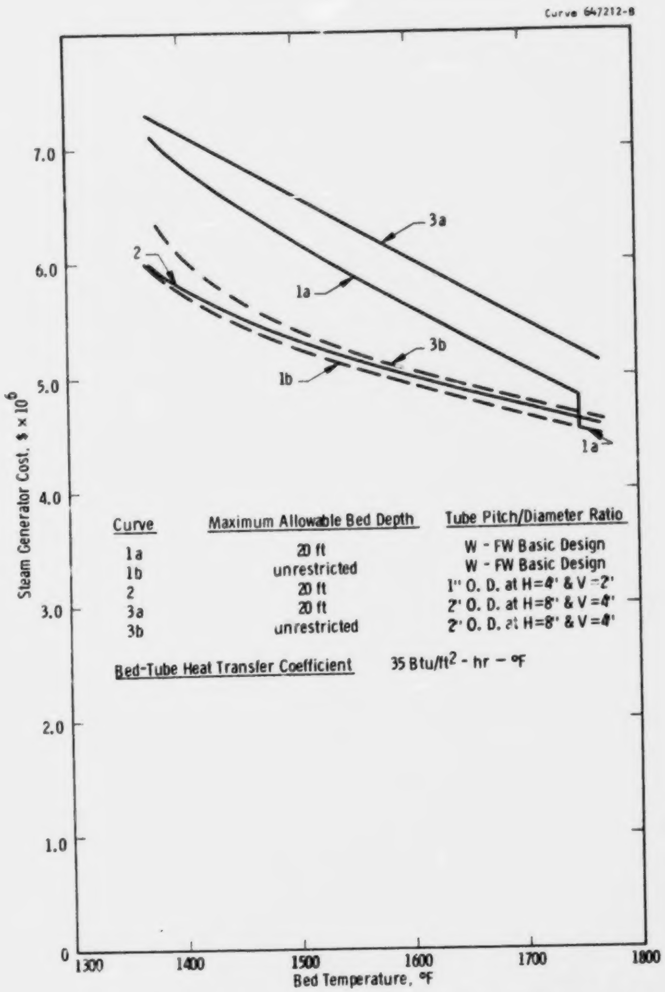


Figure 44: Effect of Tube Pitch/Diameter Ratio on the Steam Generator Cost (not including erection)

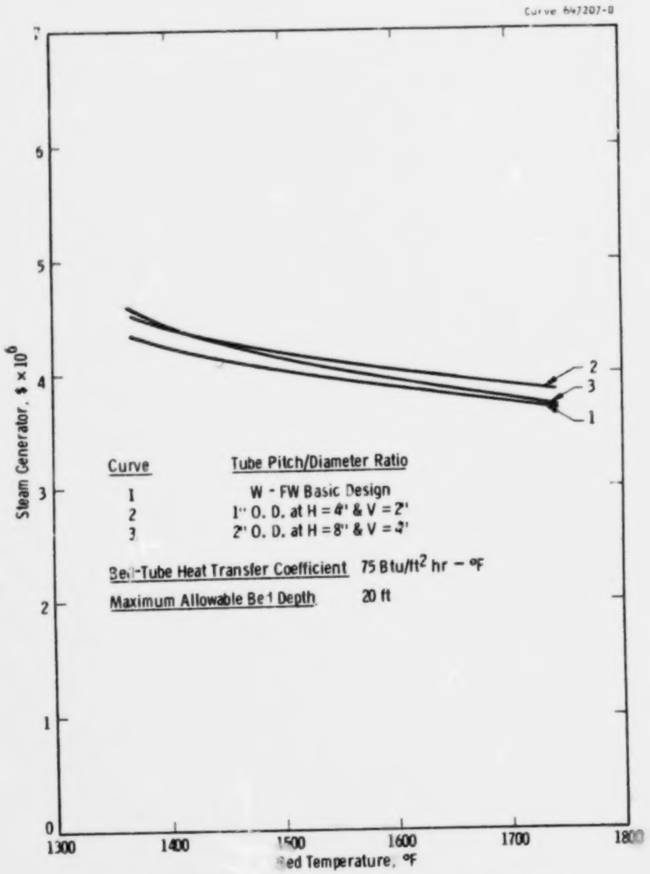


Figure 45: Effect of Tube Pitch/Diameter on the Steam Generator Cost (not including erection)

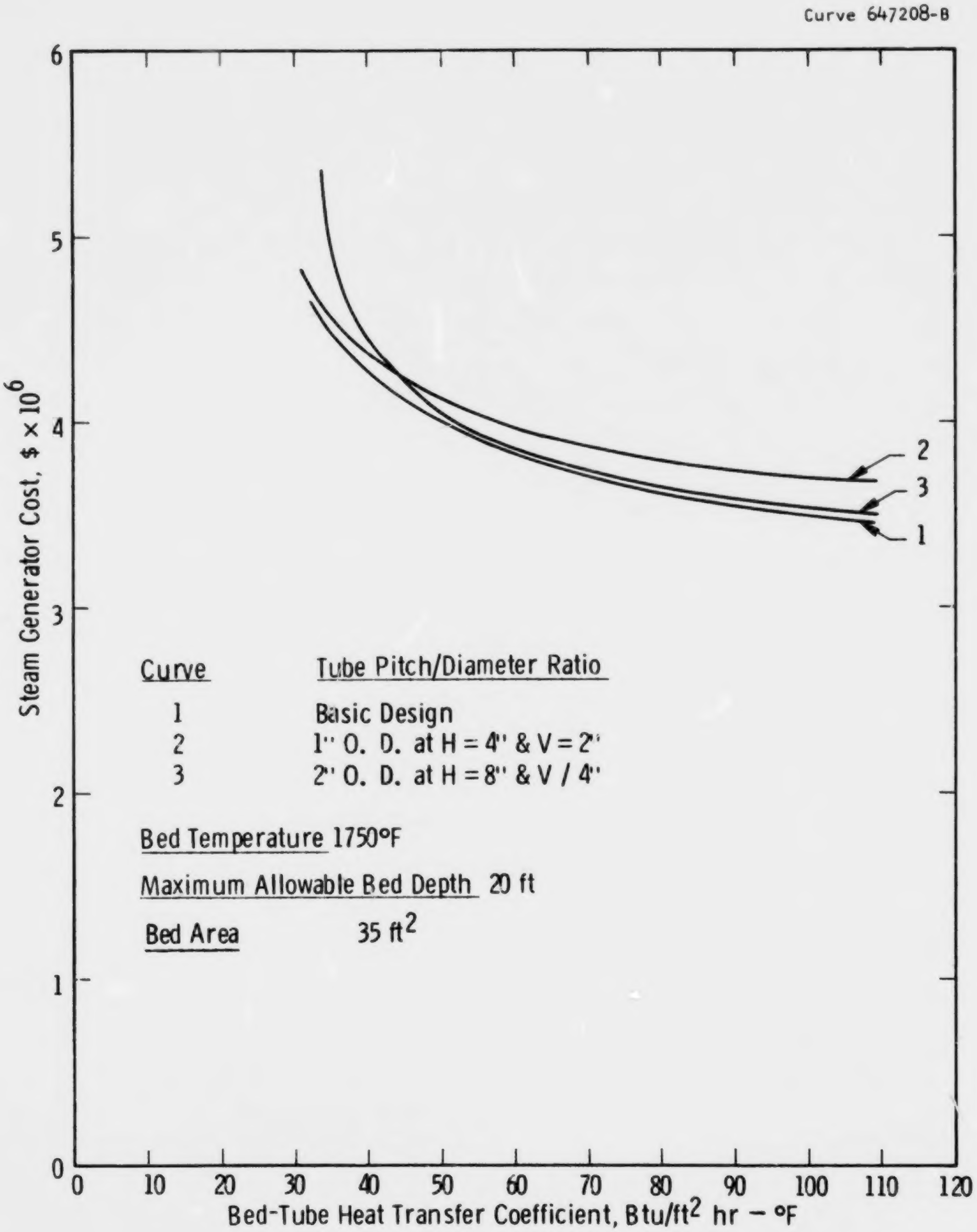


Figure 46: Effect of Bed-Tube Heat Transfer Coefficient on the Steam Generator Cost (not including erection)

a further decrease in the bed-tube heat transfer coefficient lower than 50 Btu/ft^2-hr-°F increases the steam generator cost rapidly, especially for large tube sizes and tube spacings and at lower design bed temperatures (Figures 46 and 47). A 40% increase in cost occurs when the bed-tube heat transfer coefficient decreases from 50 to 35 Btu/ft^2-hr-°F at design bed temperatures of 1636°F for 2-in.O.D. tubes at H = 8 in.and V = 4 in.(curve 3, Figure 42). It is recommended that the steam generator be designed at a bed-tube heat transfer coefficient of around 75 Btu/ft^2-hr-°F if it is at all possible and avoid designing the steam generator at a bed-tube heat transfer coefficient lower than 50 Btu-ft^2-hr-°F, especially if large tubes and spacings are used and if lower bed temperatures are employed.

To complete the evaluation, the cost information was also prepared for other cases at different design and operating variables to show the interacting effect of the tube pitch/diameter ratio, the bed-tube heat transfer coefficient, the maximum allowable bed depth, and the design bed temperature. The curves are presented in Appendix B.

Again, the maximum allowable bed depth turns out to be the limitation of the steam generator design and cost, especially at a low bed-tube heat transfer coefficient where a larger amount of heat transfer surface is required in the bed. In this case, a smaller tube size and tube spacing and a higher design bed temperature are preferred. At a maximum allowable bed depth of 10 ft, the boiler designed at 35 Btu/ft^2-hr-°F costs \$3.5/kw more than that designed at 50 Btu/ft^2-hr-°F and \$8.6/kw more than that designed at 75 Btu/ft^2-hr-°F at 1750°F design bed temperature. At a design bed temperature of 1407°F, the figures are \$6.7/kw and \$12.5/kw respectively. With 1-in.O.D. at H = 4 in. and V = 2 in. the figures are \$3.3/kw and \$4.4/kw at 1750°F; and \$2.8/kw and \$8.1/kw at 1407°F respectively.

Cost savings become smaller when the bed-tube heat transfer coefficient is further increased over 75 Btu/ft^2-hr-°F. If the bed-tube heat transfer coefficient is decreased to lower than 50 Btu/ft^2-hr-°F,

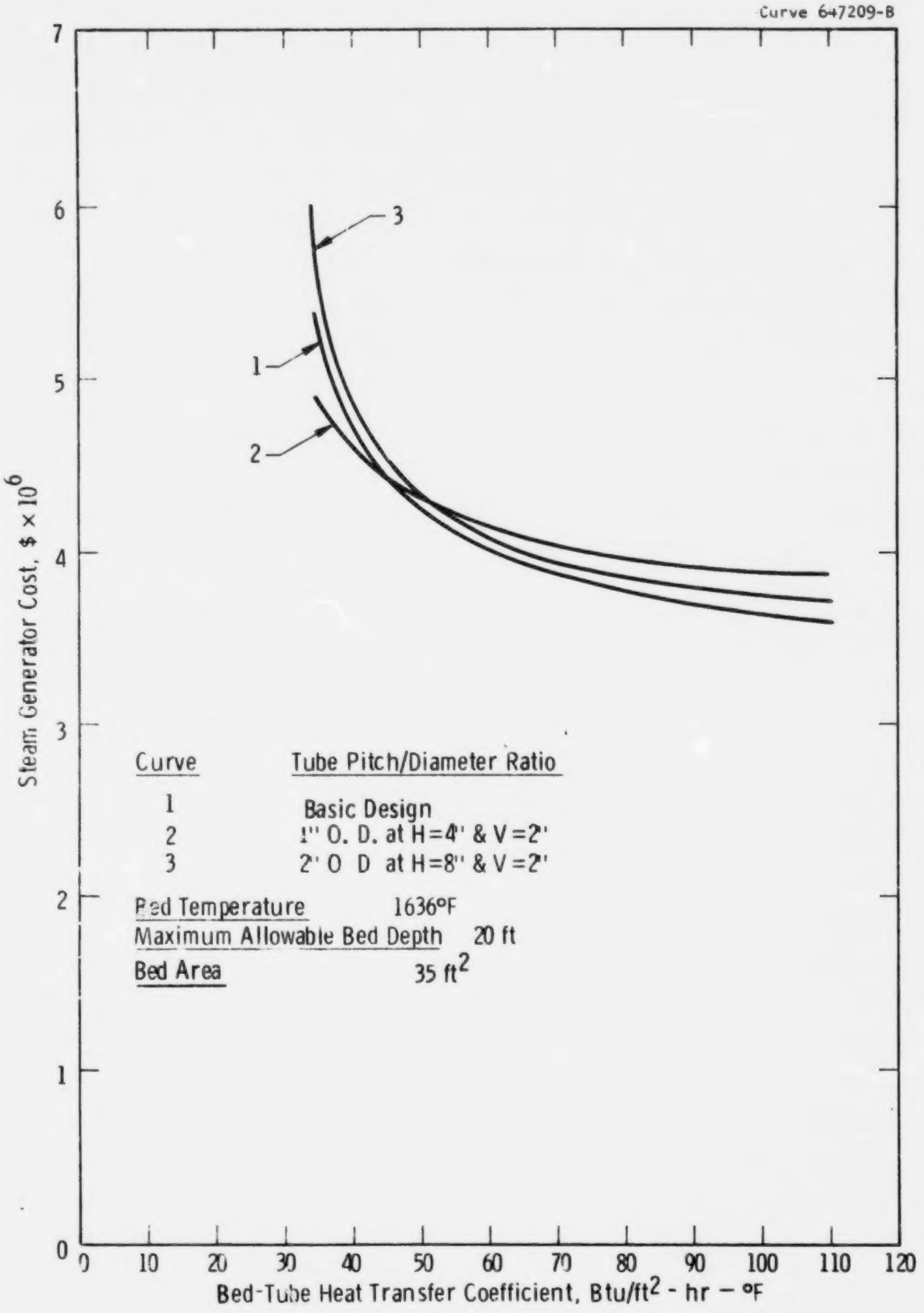


Figure 47: Effect of Bed-Tube Heat Transfer Coefficient on the Steam Generator Cost (not including erection)

the steam generator cost increases rapidly. Thus, it is recommended that the bed-tube heat transfer coefficient be kept at higher than 50 Btu/ft^2-hr-°F, and preferably around 75 Btu/ft^2-hr-°F, at the current design conditions.

2.3.3 Tube Materials

The basic design specifies conventional boiler tube material, with SA-210-A1 for tubes in the preevaporator; SA-213-T2 for tubes in the lower superheater; SA-213-T22 for tubes in the water walls, upper superheater (lower loops), and reheater; and SA-213-TP304H for tubes in the upper superheater (upper loops). Changes in design bed temperature, bed-tube heat transfer coefficient, or steam temperature may require higher grade tube materials. This would not, however, substantially affect the steam generator cost, because the tubing cost alone constitutes only $\sim$ 10% of the total steam generator cost ($\sim$ \$1.40/kw). Higher fabrication cost for higher alloy material may increase this cost slightly. Nevertheless, the total boiler cost is not expected to increase significantly because of a change of tube material, unless the operating bed temperature and heat transfer coefficient are drastically changed. The selection of tube materials at different design bed temperatures and bed-tube transfer coefficients is presented in Appendix A.

2.3.4 Module Capacity

Boiler modules can be shop fabricated, partially shop fabricated, or field erected, depending on the size. Modules up to 12 ft in diameter can be shop fabricated; those up to 17 ft in diameter can be partially shop fabricated (the boiler internals shop fabricated, the pressure shell field erected). The plant concept for a given capacity can be either multiples of shop-fabricated modules, partially shop-fabricated modules, or field-erected modules.

The boiler plant equipment cost is different for each case. The steam generator cost will depend on operating conditions and design variables such as design bed temperature, tube size and tube spacing, maximum allowable bed depth, etc. Auxiliary equipment will also be affected: coal feeding, limestone or dolomite feed and withdrawal, particulate removal, steam piping, boiler feedwater system, etc.

The evaluation of these approaches is based on Figure 8 where the cost variation of the pressure parts, shell, subcontracted and contracted equipment, drafting and home office, and erection with respect to the plant size is presented. The resulting costs (including erection) are presented in Figure 48 for a design bed temperature of 1750°F, and in Figure 49 for a design bed temperature of 1636°F at a maximum allowable bed depth of 20 ft. The results show that at a plant size larger than ∿ 340 MW, partially shop-fabricated four-module plants with maximum shop fabrication of the pressure parts are more economical than a collective multiple of the largest shop-fabricated modules of the same plant capacity. Changing the tube size and tube spacing does not affect this conclusion (Figures 48 and 49). A change in heat transfer coefficient should not produce a different conclusion because the cost escalation due to the addition of a single module is greater than it would be if the existing modules were simply enlarged. However, this is no longer true when the shell size is larger than ∿ 17 ft, because the degree of shop fabrication of the pressure parts decreases and the steam-generator cost again increases at a much higher rate.

When the maximum allowable bed depth is decreased from 20 ft, the cost saving of the four-module partially shop-fabricated plant is expected to increase. The splitting of the beds in the shop-fabricated module means an increase in module height. In the partially shop-fabricated module, however, the splitting of beds can be avoided simply by enlarging the module diameter. From the discussion on Figure 19, the latter means a more economical alternative except when the module diameter is increased beyond ∿ 17 ft.

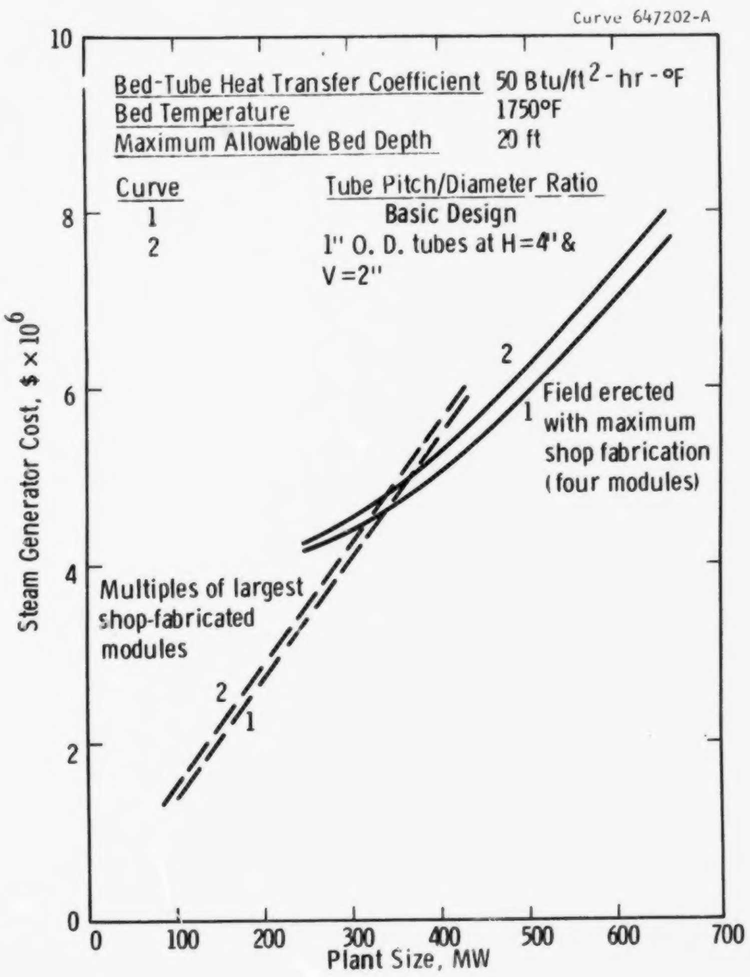


Figure 48: Dependence of the Steam Generator Cost on the Plant Size

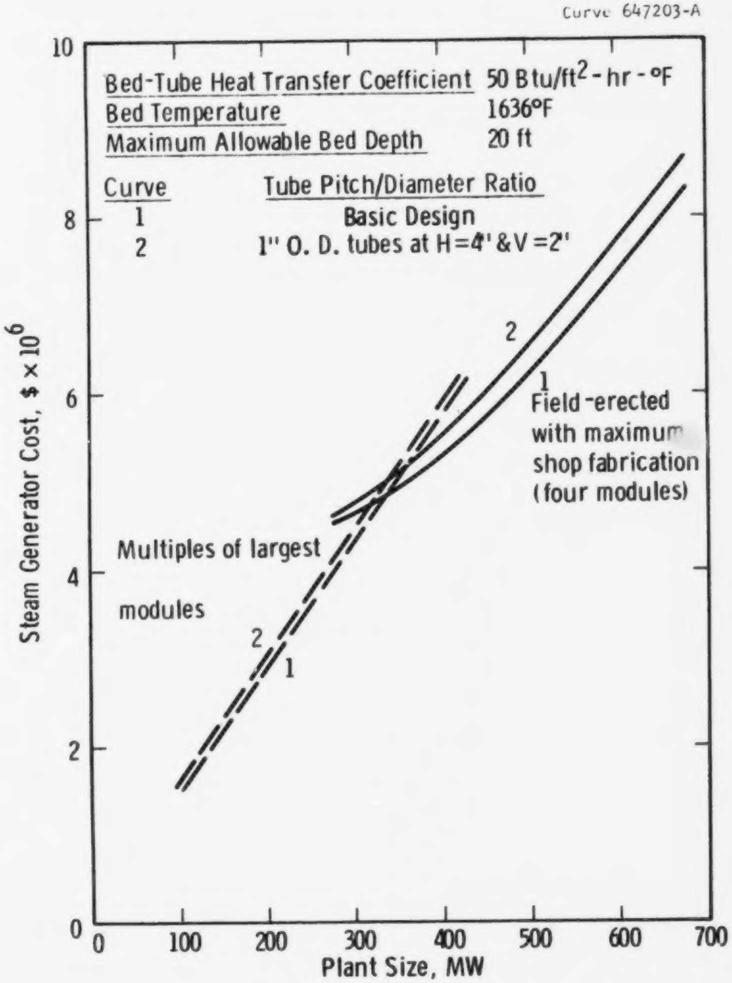


Figure 49: Dependence of the Steam Generator Cost on the Plant Size

The optimum design variables (fluidizing velocity, excess air and pressure) in relation to module capacity are discussed in their respective sections.

2.4 PARTICULATE REMOVAL SYSTEM ECONOMICS

The primary variables taken into consideration in analyzing particulate removal system economics are dust loading, particle size distribution entering and leaving the system, and gas flow rate. The effects of boiler operating variables -- bed temperature, freeboard height, and superficial velocity; and design variables -- tube pitch/diameter ratio, and bed-tube heat transfer coefficient were also evaluated.

2.4.1 Range of Dust Loadings and Particle Size Distribution

The cases evaluated are outlined in Table 10. Due to the complexity of boiler design and operating parameters, a given dust loading, such as one identical to that of the basic design, can be produced by infinite combinations of these parameters (Figure 50). The basic design is shown as a point which represents an 8.5% ash coal with 6% carbon elutriation rate and 0.7% dolomite elutriation rate. For the same coal, a 0% carbon elutriation rate will require a 5.3% dolomite elutriation rate to produce the same dust loading. For a coal of 13.4% ash, the elutriated ash alone will give the same dust loading with no carbon or dolomite elutriated. Similar diagrams are also prepared for the cases which double and triple the basic design dust loading (Figures 52 and 53). The results of this parametric study are presented in Appendix C.

2.4.2 Gas-Turbine Specification

A review of operating experience and an assessment of erosion in gas turbines was prepared by Westinghouse under contract to the Office

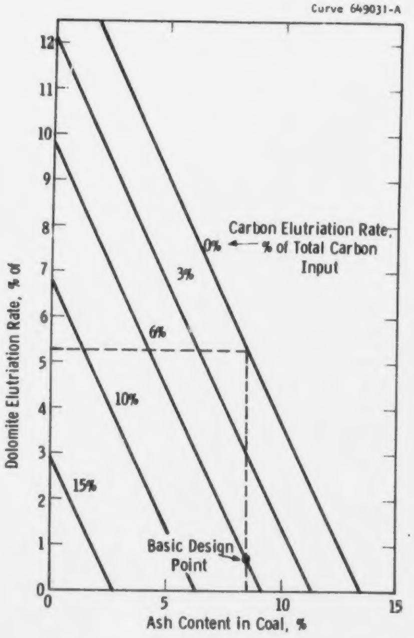


Figure 50: Different Combinations of Elutriation Rates for a Dust Loading Equal to That of Ⓦ-FW Basic Design

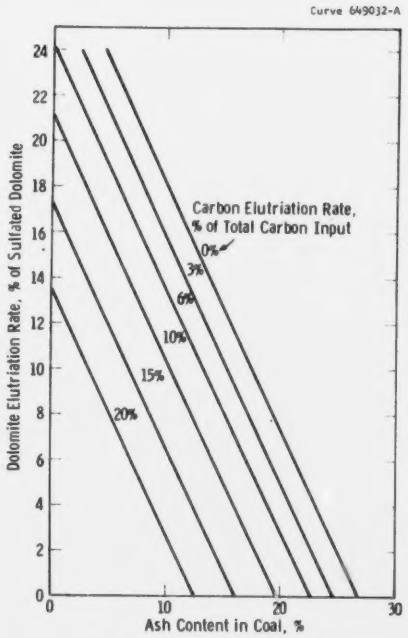


Figure 51: Different Combinations of Elutriation Rates for a Dust Loading Double That of Ⓦ -FW Basic Design

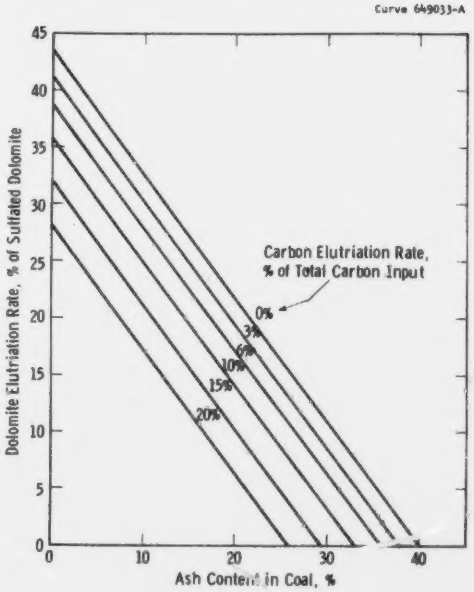


Figure 52:

Different Combinations of Elutraition Rates for a Dust Loading Triple That of Ⓦ-FW Basic Design

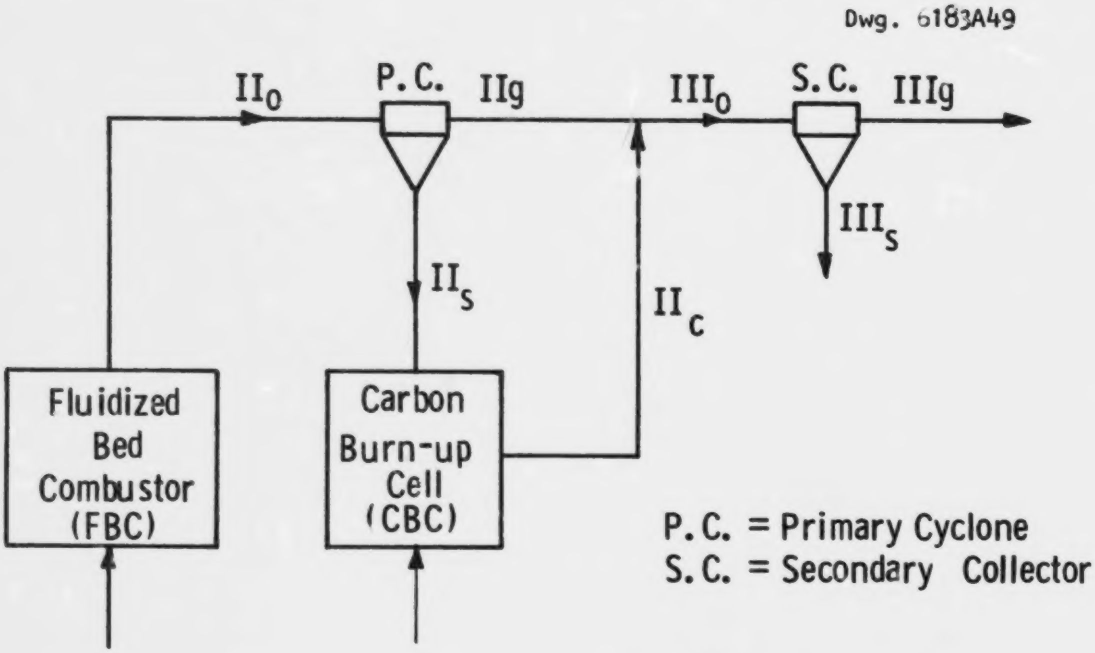

Dwg. 6183A49
II_0
P.C.
IIg
III_0
S.C.
IIIg
II_s
III_s
II_c
Fluidized Bed Combustor (FBC)
Carbon Burn-up Cell (CBC)
P.C. = Primary Cyclone
S.C. = Secondary Collector

Figure 53: Flow Diagram for Particulate Removal System

TABLE 10

CASES EVALUATED FOR DETERMINING EFFECT OF DUST LOADING AND PARTICLE SIZE DISTRIBUTION

Cases Evaluated	Dust Loading Leaving FBC	Particle Size Distribution	Cyclone System Design
Group 1 - Case 1	basic design (6.7 grains/SCF)	(refer to Figure 56)	Figure 53
Group 1 - Case 2	double design value	curve 1 for particles elutriated from FBC	Figure 53
Group 1 - Case 3	triple design value	curve 2 for particles elutriated from CBC	Figure 53
Group 1 - Case 4	triple design value		Figure 54
Group 1 - Case 5	triple design value		Figure 55
Group 2 - Case 1	basic design (6.7 grains/SCF)	curve 2 for particles elutriated from FBC	Figure 53
Group 2 - Case 2	double design value	curve 3 for particles elutriated from CBC	Figure 53
Group 2 - Case 3	triple design value		Figure 53

of Air Programs.[1] Specifications for the fluidized bed combustion system based on that study are:

- Dust loading less than 0.15 grains per standard cubic foot
- Concentration of particles greater than 2 micrometers less than 0.01 grains per standard cubic foot.

These design requirements will be updated as additional laboratory test data and operating experience become available.

2.4.3 Effect of Dust Loading and Particle Size Distribution on Cost

The effect of dust loading and particle size distribution leaving the boiler on the particulates going to the gas turbine for the particle removal systems selected is shown in Table 11. The gas-turbine specification is exceeded in several cases. In order to meet the specification, alternative particulate removal systems could be considered -- granular bed filters, electrostatic precipitators, ceramic filters, etc.; additional mechanical collectors could be used in series; or the boiler operation could be altered. The additional use of mechanical collectors was the approach selected because the equipment is available and thus provides the best cost data and because the effect of boiler operating conditions on particulate emission is difficult to project. A summary of the economic implications is presented in Table 12.

If the gas-turbine dust loading requirement is defined as < 0.01 grains/SCF without reference to particle size, no centrifugal separator presently available can meet this requirement within reasonable cost in all cases discussed above. If such separators were available, high temperature ceramic filters might have to be used. The possibility of this application is being evaluated. In this respect, the particle size distribution is a far more important parameter than the dust loading, since the collection efficiency for the particle size smaller than 2μ

TABLE 11

DUST LOADINGS LEAVING THE SECONDARY CYCLONE FOR DIFFERENT CASES EVALUATED

Cases Evaluated	Total Dust Loading Leaving the Secondary Cyclone, gr/SCF	Dust Loading for Particles Larger than 2 μ, gr/SCF
Group 1 - Case 1	0.14	0.007
Group 1 - Case 2	0.27	0.014
Group 1 - Case 3	0.40	0.020
Group 1 - Case 4[a]	0.16	0.0006
Group 1 - Case 5	0.40	0.019
Group 2 - Case 1	0.27	0.013
Group 2 - Case 2	0.46	0.022
Group 2 - Case 3	0.67	0.032

[a] Assuming same fractional collection efficiency for the second secondary cyclone as that for the first one. This assumption is too optimistic.

TABLE 12

PARTICULATE REMOVAL SYSTEM COST

Case	Need for Further Clean-up to Achieve Gas-Turbine Specification	Particulate Removal System Cost Increase $/kw
Group 1 - Case 1	no	--
Group 1 - Case 2	may not be required[a]	--
Group 1 - Case 3	yes, a second secondary cyclone in series or granular bed filter	6.30 8.50
Group 1 - Case 4	no	--
Group 1 - Case 5	yes, a second secondary cyclone in series or granular bed filter	6.30 8.50
Group 2 - Case 1	may not be required[a]	--
Group 2 - Case 2	yes, a second secondary cyclone in series or granular bed filter	6.30 8.50
Group 2 - Case 3	yes, a second secondary cyclone in series or granular bed filter	6.30 8.50

[a] Dust loading is larger than specification but particle size is close to specification.

decreases rapidly. Thus, the particle size distribution curves assumed for the present evaluation (as shown in Figure 56) are conservative because a large amount of fines is assumed to be present.

2.4.4 Effect of Gas Flow Rate on Cost

The effect of increasing the gas flow rate was evaluated for four cases which correspond to the basic design flow rate, 30% excess air, 50% excess air, and 100% excess air. The selection of first-stage cyclones is based on the criterion of maximum efficiency at minimum cost with a minimum cyclone efficiency of 85% (Table 13). The cost increment for a higher gas flow rate is shown in Figure 57.

The first-stage cyclone cost includes not only the cost of the first-stage separators supplied by Ducon but also the cost of the separator pressure vessel and all of the gas piping from the steam generator outlet to the secondary separator inlets. The costs for the pressure vessel and gas piping from the steam generator to the first-stage separator is lined with hard refractory, but the pressure vessel and the gas piping from it are lined with stainless steel. In the other case, hard refractory without an alloy liner was used throughout. At 100% excess air, the increase is $1.70/kw and $2.40/kw for a hard refractory liner and a stainless steel liner, respectively. The major cost increase is from the enlargement of pressure vessels and gas piping due to the higher gas flow rate. The increase in separator cost alone constitutes only about 18% of the total cost increment.

The selection of the secondary collectors is shown in Table 14, and the cost increase for the second stage is presented in Table 58. The incremental cost for the second stage is more than twice that of the first stage ($5.00/kw versus $2.40/kw at 100% excess air). This is because model 18,000 is the largest cyclone now supplied by Aerodyne. The capacity of model 18,000 with dirty gas as the secondary gas is 30,000 CFM. Any gas flow rate higher than that will require multiple units with their individual pressure vessel. This would tend to increase

TABLE 13

SELECTION OF FIRST-STAGE CYCLONES AT DIFFERENT GAS FLOW RATE

Operating Condition	Basic Design	30% Excess Air	50% Excess Air	100% Excess Air
Actual Gas Flow at First Stage, ACFM per module	59,000	71,000	80,000	106,000
Cyclone Selection, Duclone	size 355 model 810	size 440 model 810	size 500 model 810	size 635 model 810
Number of Cyclones per Module	4	4	4	4
Pressure Drop, in W.G.	19.7	18.6	19.2	20.9
Collection Efficiency, % by Weight[a]	89.5	85.5	85.3	85.1

[a] Collection efficiency based on particle size distribution of Group 1.

TABLE 14

SELECTION OF SECOND-STAGE CYCLONES AT DIFFERENT GAS FLOW RATE

Operating Condition	Basic Design	30% Excess Air	50% Excess Air	100% Excess Air
Actual Gas Flow at First-Stage Conditions, ACFM per Module	59,000	71,000	80,000	106,000
Cyclone Selection, Aerodyne Model Number	18,000	15,000	18,000	15,000
Number of Cyclones per Module	2	3	3	4
Pressure Drop, in W.G.	31	31	31	31
Collection Efficiency, % by Weight[a]	97	97	97	97

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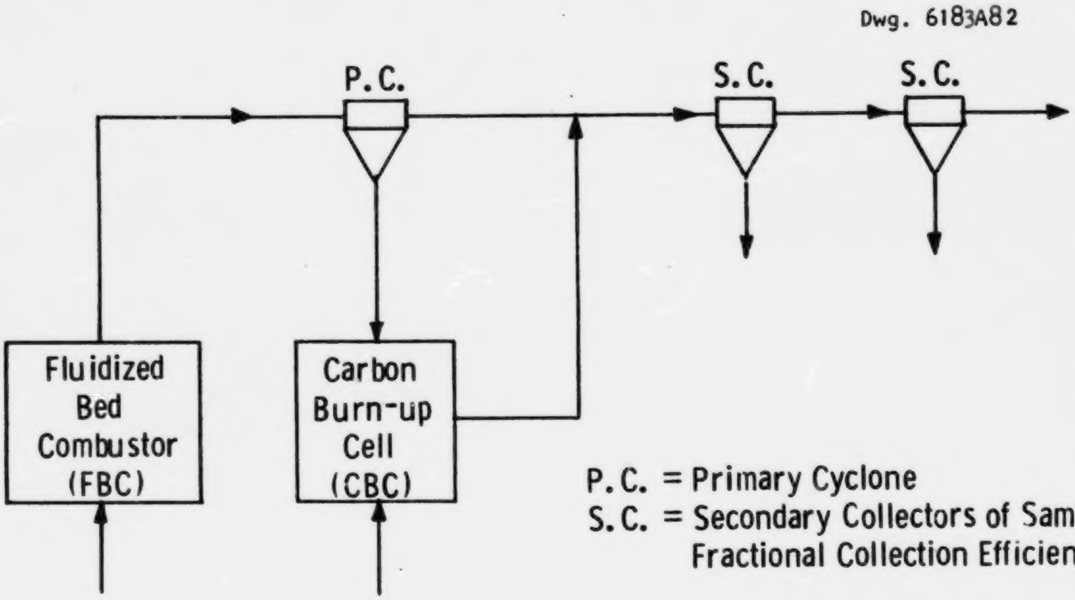
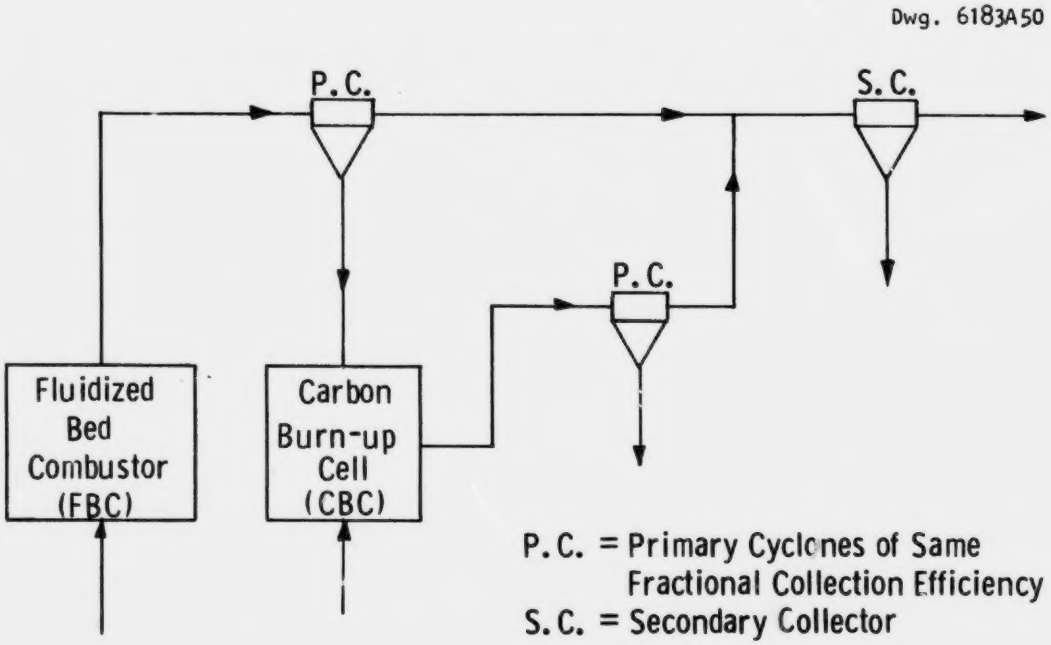
[a] Collection efficiency based on particle size distribution of Group 1.


Dwg. 6183A82
P.C.
S.C.
S.C.
Fluidized Bed Combustor (FBC)
Carbon Burn-up Cell (CBC)

P.C. = Primary Cyclone
S.C. = Secondary Collectors of Same Fractional Collection Efficienc

Figure 54: Flow Chart for Group 1 - Case 4


Dwg. 6183A50
P.C.
S.C.
P.C.
Fluidized Bed Combustor (FBC)
Carbon Burn-up Cell (CBC)

P.C. = Primary Cyclones of Same Fractional Collection Efficiency
S.C. = Secondary Collector

Figure 55: Flow Chart for Group 1 - Case 5

Curve 649047-A

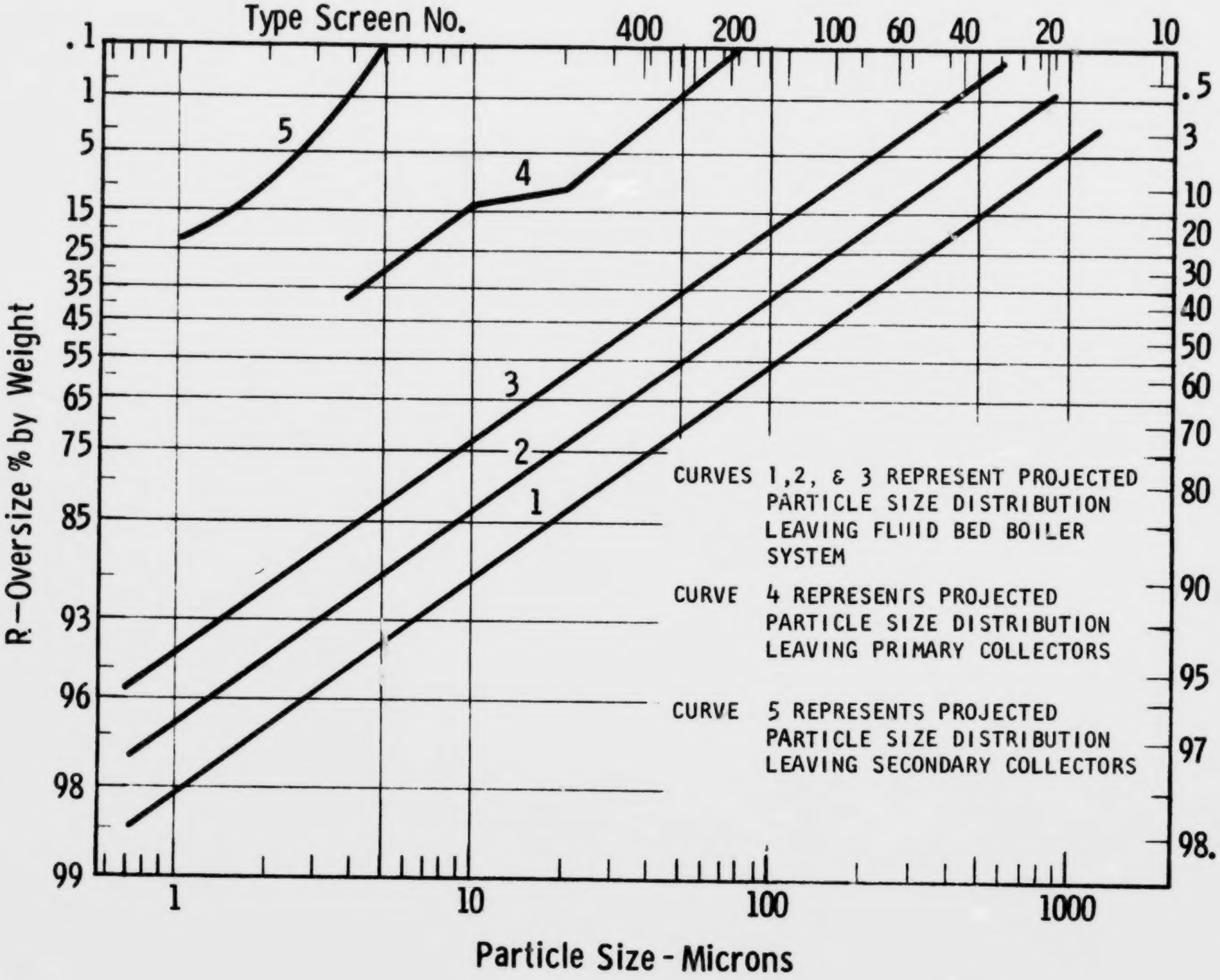


Figure 56: Particle Size Distribution for Different Gas Streams

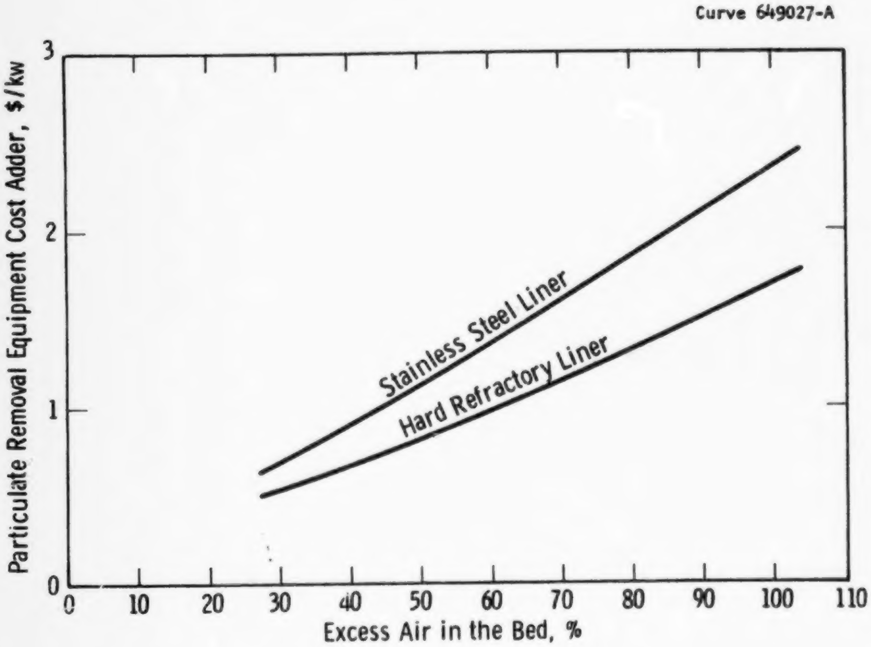


Figure 57: Increase in First-Stage Cyclone Cost Due to Increase in Gas Flow Rate (for 318 MW Plant)

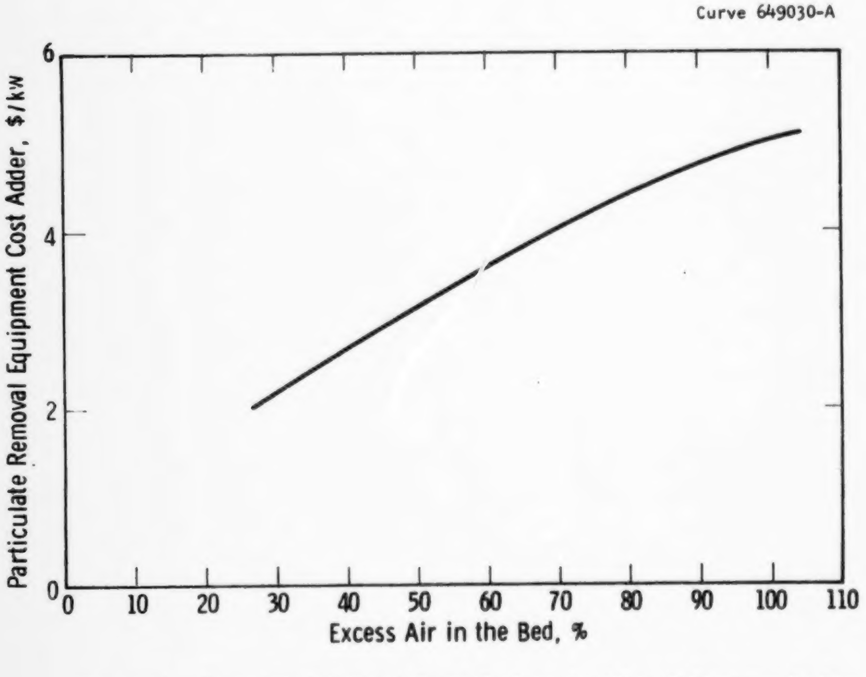


Figure 58: Increase in Second-Stage Cyclone Cost Due to Increase in Gas Flow Rate (for 318 MW Plant)

the incremental cost of the second stage, but the rate of cost increase slows down at excess air larger than 100%. At excess air larger than 100%, the rate of cost increase for the first stage speeds up. This is because increases in pressure vessel size and gas piping diameter increase the incremental cost rapidly at very high gas flow rates.

Taking into account the cost of gas piping from the second-stage cyclones to the gas turbine, the total incremental cost at different gas flow rates is shown in Figure 59.

Combining cost figures from Figures 26, 27, and 59, an increase in total boiler cost of $9.6/kw to $10.2/kw is required to operate the boiler at 100% excess air. This, however, does not take into consideration the fact that the combustion efficiency in the primary beds will approach 100% at 100% excess air and that the carbon burn-up cell can be eliminated. In addition, the particulate removal system would be much simpler and the high excess air might also provide the necessary flexibility to achieve plant turndown if the boiler were designed at lower bed temperatures. An additional ∿ 10% turndown capability is obtained by operating at 100% excess air. The lower bed depths would also permit combining the two superheater beds into one bed which would also reduce module height and cost. Thus, it is concluded that if the carbon burn-up cell can be eliminated, operating at 100% excess air may result in a lower energy cost, with increased cycle efficiency and flexibility. This conclusion is true at a 300 MW nominal plant size but is not necessarily true at a 600 MW plant size. This is because at the 600 MW plant size, the module is 17 ft in diameter in the basic design. A further increase in excess air requires either an increase in module diameter or in the total number of modules. At 100% excess air, four modules of 23 ft in diameter are required. Since the fabrication cost of the internals increases rapidly at module diameters larger than 17 ft because of the rapidly decreasing portion which can be shop-fabricated, an increase of $5.0/kw in boiler module cost alone is conceivable. Adding the cost increase in the particulate removal system and in turbine equipment, the total increase in boiler cost is about $15.0/kw. In this case, operating at higher pressures may be beneficial.

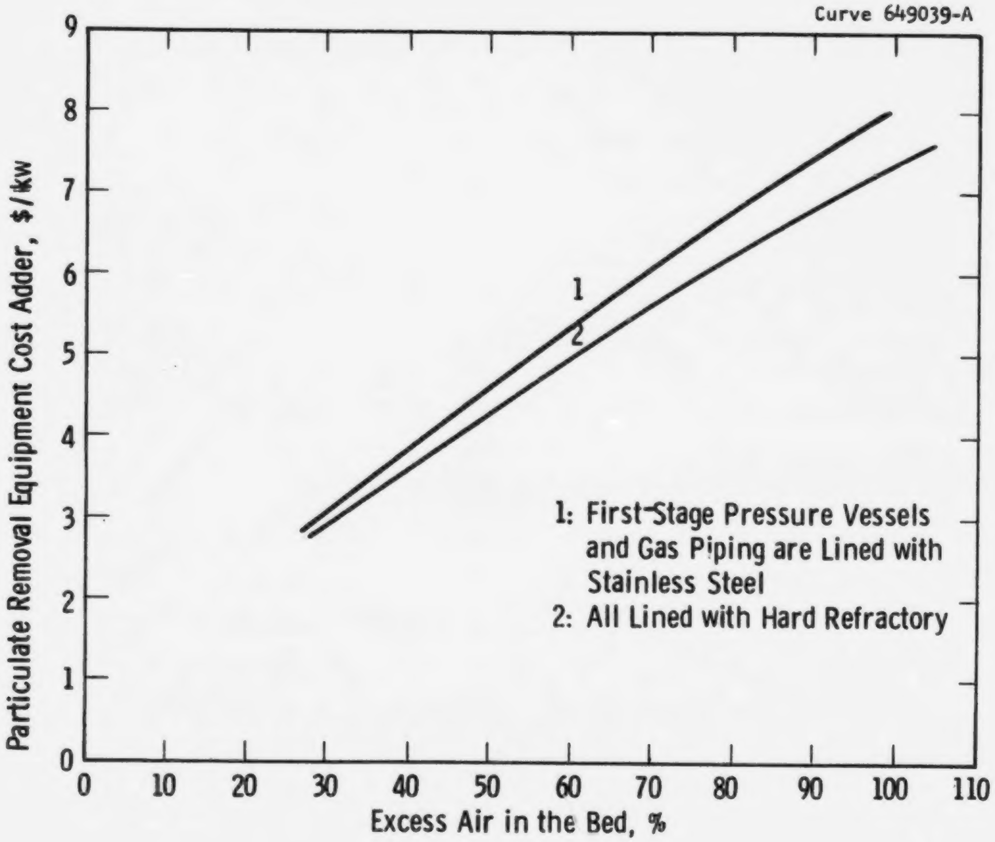


Figure 59: Cost Increments for Particulate Removal System at Different Gas Flow Rates

Instead of increasing the module diameter from 17 ft to 23 ft, the number of modules could be increased but the module diameter kept constant. At 100% excess air, seven modules are required. In this case, in addition to module cost, individual particle removal equipment has to be provided for each module; ducting and piping manifolds have to be increased; the coal feeding system becomes more complicated; and, above all, instrumentation and control have to be more sophisticated. The total increase in boiler cost is estimated to be $20.0/kw to $25.0/kw in this case. Consequently, even if the carbon burn-up cell could be eliminated, the increase in cost and complication in control do not clearly favor the operation at high excess air for large plant size. A careful evaluation of overall design and control philosophy should be done if operation at high excess air is to be attempted for large plant size.

2.5 POWER PLANT

Alternative boiler design and operating conditions will have three primary affects on the power generation equipment:

- Capacity of gas- and steam-turbine equipment
- Gas-turbine inlet temperature
- Ability to achieve higher steam temperatures and pressures.

Since this analysis assumes a constant fuel rate, the capacity changes are considered small except for the high excess air case. The effect of capacity was considered with the excess air analysis.

2.5.1 Gas-Turbine Inlet Temperature

The design value for the gas-turbine inlet temperature was 1600°F for the base design of the pressurized fluid bed boiler. This is well below the current state-of-the-art temperatures of 1800 to 1900°F for utility intermediate load applications. The 1600°F level was established by assuming the flue gas leaves the bed at the 1750°F bed temperature and that the temperature difference between the bed and the gas-turbine inlet would be 150°F. Several factors may alter the gas-turbine inlet temperature. These include:

- Temperature drop between boiler and turbine expander: 150°F was assumed but is probably excessive. A drop as low as 50°F to 75°F may be achieved.
- Bed temperature. If the boiler design temperature is changed, the turbine inlet temperature will change. Sulfur removal considerations and ash agglomeration will determine the maximum temperature.
- Combustion above the bed. Combustion has been observed above the bed of a fluidized bed boiler which increases the gas temperature 200 to 300°F. Any combustion above the bed would increase the gas-turbine inlet temperature. No combustion was assumed in the base design.
- Modification of the cycle to provide for reheat of the product gas from the boiler prior to the gas turbine. One concept for doing this is shown in Figure 60. Carbon carried out of the primary beds would be gasified to produce a low-Btu gas. The gas would be used in the second-stage combustor.

Performance calculations were made to determine the effect of a change in the gas-turbine inlet temperature on plant performance. The results are summarized in Table 15. These results are for 17.5% excess air and a boiler efficiency of 88.6%. A 200°F change in the turbine inlet temperature will change the plant heat rate $\sim$ 1% using current technology.

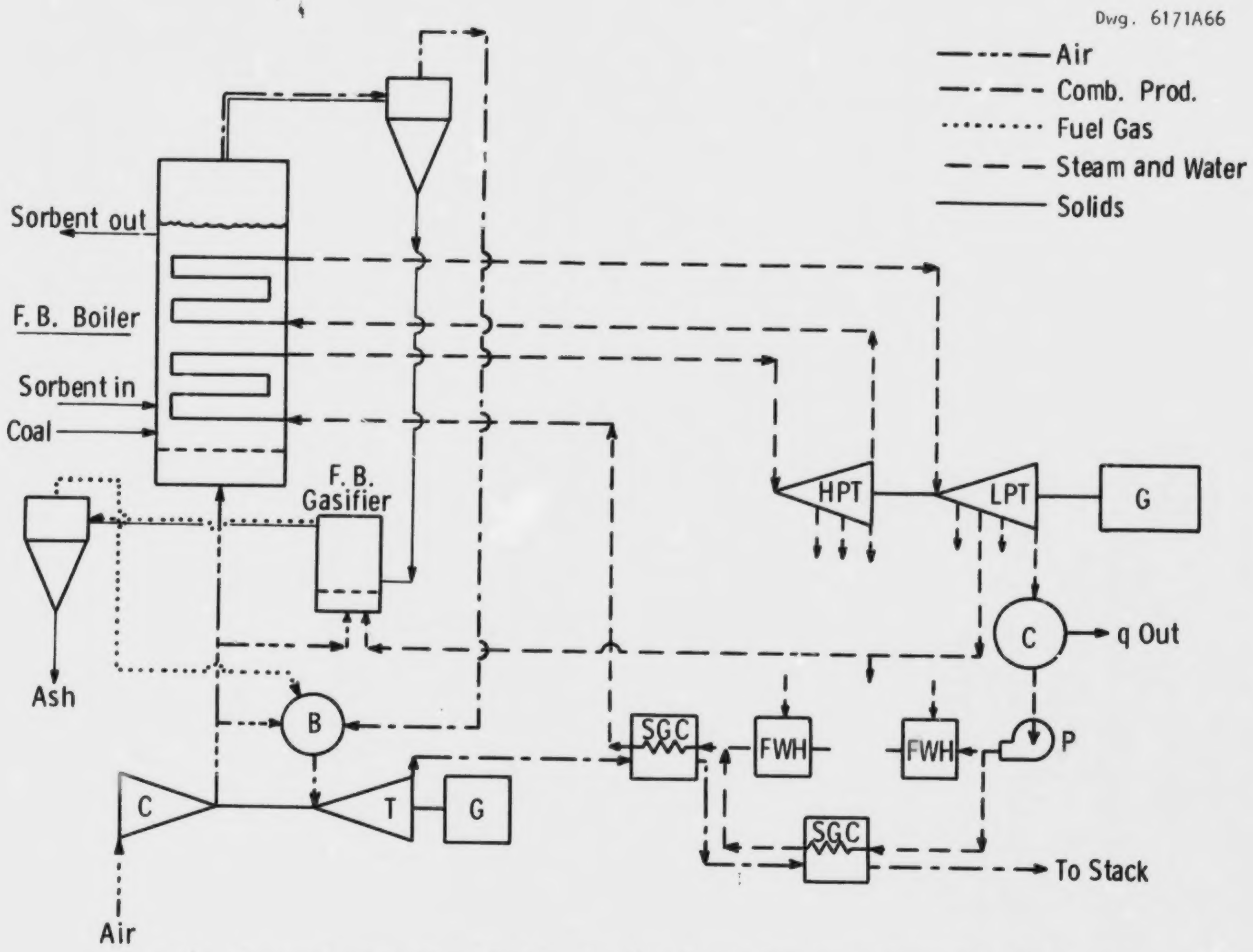

Dwg. 6171A66
Air
Comb. Prod.
Fuel Gas
Steam and Water
Solids
Sorbent out
F. B. Boiler
Sorbent in
Coal
F. B. Gasifier
Ash
B
C
T
G
HPT
LPT
G
C
q Out
P
SGC
FWH
FWH
SGC
To Stack
Air

Figure 60: Pressurized Fluidized Bed Combustion Power Plant with Secondary Combustor

TABLE 15

PERFORMANCE OF PRESSURIZED FLUID BED BOILER POWER PLANT AS A FUNCTION OF GAS-TURBINE INLET TEMPERATURE

Gas-Turbine Inlet Temperature °F	Plant Output[a] MW	Plant Heat Rate[a] Btu/kWh	Fuel Burned after Primary Beds[b] %
1400	625.6	9293	--
1500	634.9	9157	--
1600	644.1	9026	--
1700	644.8	8921	2.5
1800	645.3	8820	5.0
1900	641.8	8773	7.6
2000	636.2	8753	10.1

[a] The decrease in performance at high gas-turbine inlet temperatures is the result of increased bleed air required for turbine cooling and the increase in gas-turbine waste heat which reduces steam cycle extraction for regenerative heating. The heat rate at 2000°F would be reduced ∿ 270 Btu/kWh if turbine blade cooling was not required.

[b] Assumes any increase above the design value of 1600°F has to result from burning fuel either above the bed or separately.

2.5.2 Steam Temperature

Plant performance can be increased by increasing steam temperature and pressure. The effect of higher steam temperatures on the performance of the plant is shown in the following:

Steam Temperature °F	Steam Pressure psi	Gas-Turbine Inlet Temperature °F	Power MW	Heat Rate Btu/kWh
1000/1000	2400	1600	644.1	9026
1100/1100	3300	1600	673.9	8627
1200/1200	4500	1600	690.8	8417

An increase of 100°F in both superheat and reheat temperatures will give a reduction of about 400 Btu/kWh in plant heat rate. The increased performance and inherently less severe boiler tube corrosion in fluidized bed boilers make high steam temperatures attractive.

2.6 ASSESSMENT

A summary of the sensitivity analysis is presented in Table 16. Each parameter is indicated with a projection of what change might be required as the result of experimental data. For example, the bed temperature was set at 1750°F for 100% load. This temperature may prove to be too high for economical sulfur removal and have to be reduced to 1600 or 1650°F, which may be more favorable. The summary table indicates what the effect of such a change would be on plant cost and performance, assuming no other variable restrictions. In this case, the plant cost would increase < \$3/kw, plant performance would decrease < 0.5%, and plant turndown to 25% load could still be achieved. This would result in an energy cost reduction of < 0.3 mills/kWh. Since the projected advantage over a conventional plant with stack gas scrubbing is greater than 1.5 mills/kWh,[1] the penalty for lowering the temperature is not significant. This conclusion holds for all of the variables considered and is generally valid for plant capacities of 200 to 700 MW. The results of the sensitivity analysis indicate that the base plant design is relatively insensitive to changes in operating conditions or design parameters.

TABLE 16

SENSITIVITY ANALYSIS SUMMARY

		Effect			
		Plant Cost			
Parameter	Change	Boiler	Auxiliaries	Power Generation Equipment	Performance
Boiler Operating Conditions					
Bed Temperature	Reduction from 1750°F to 1600°F	< $3/kw(1) increase	negligible	negligible	• ≤ 0.5% decrease in efficiency • 4:1 turndown requirement can be met
Fluidizing Velocity	decrease to 5 fps increase to 15 fps	< $0.50/kw increase ~ $1/kw increase(3)	(2) (2)	- -	- -
Excess Air	increase to 100%	< $1/kw decrease(4)	~ $7.50/kw increase	< $4/kw increase(5)	• ~ 0.5% increase in plant efficiency • improved turndown capability
Pressure	increase to 15 ata	< $0.50/kw increase for constant area ~ $1/kw increase for constant velocity	$4/kw decrease	negligible	~ 0.2% decrease in plant efficiency
Particulate Carry-Over					
Loading	increase loading to 3 times the design value (> 20 gr/SCF)	-	$6-8/kw increase(6)	-	negligible
Particle Size	increase fines: particles < 10 μm increased from 15% to 25%	-	no increase projected(6)	-	negligible
Boiler Design					
Heat Transfer Surface					
Configuration	increase available heat transfer surface per unit volume from 5.5 ft^2/ft^3 to 11 ft^2/ft^3	~ $1/kw increase(7)	-	-	-
	decrease from 5.5 ft^2/ft^3 to 3 ft^2/ft^3	~ $1/kw increase(7)	-	-	-
Heat Transfer Coefficient	increase from 50 to 75 Btu/hr-ft^2-°F	~ $1/kw decrease	-	-	-
	decrease from 50 to 35 Btu/hr-ft^2-°F	~ $3/kw increase	-	-	-
Materials	assume tubing cost 50% greater than base design	< $2/kw increase(8)	-	-	-
Bed Depth(9)	reduced to 10 ft	< $3/kw increase	-	-	-
Power Generation Equipment 11					
Gas Turbine Inlet Temperature	± 200°F from 1600°F	-(10)	-(10)	-(10)	± 1% in efficiency
Steam Temperature	100°F increase in superheat and reheat	< $2/kw increase	-	negligible	~ 2% increase in efficiency

---

(1) Increase will depend on bed depth restrictions; $3/kw would correspond to a maximum allowable ht of ~15 ft.
(2) The fluidizing velocity will affect the particulate removal equipment -- see particulate emission parameter for costs.
(3) Considerable savings may be realized for large capacity (> 600 mw) plants since higher velocities avoid the need for field erection.
(4) Additional savings would be realized for large capacity (> 600 mw) plants in order to avoid field erection. The $1/kw does not include a cost reduction which may result from elimination of the carbon burn up bed due to increased efficiency.
(5) $4/kw assumes the larger capacity gas turbine has the same unit cost as the base machine. Actual cost $1/kw of a larger machine would be lower.
(6) Based on projected gas turbine requirement of < 0.01 gr/SCF of particles > 2 μm.
(7) Assumes maximum allowable bed depth of 20 ft.
(8) Includes effect on fabrication.
(9) Assumes constant freeboard. (Change in freeboard requirement would have similar effect.)
(10) If the temperature is the result of equipment modification, the capital cost would be altered. Gas piping will be affected in any case.
(11) Does not include cooling tower cost which will be reduced @ higher efficiency. Cost effects are based on base design capacity and efficiency.

Operating conditions and design parameter changes can occur, most likely as the result of additive problems, which would result in a significant cost increase for the system. For example, if the bed temperature had to be decreased to 1600°F, the heat transfer coefficient was only 35 $Btu/hr\text{-}ft^2\text{-}°F$, and the dust loading from the boiler was three times the design value, the plant cost could increase by 7 to 8% and the efficiency decrease by 0.5%. This would result in an increase in the energy cost of ∿ 0.7 mills/kWh. This is a significant increase but still within the economic margin. Caution must be exercised in interpreting multiple changes in the variables. In the case above, the cost effects were added. However, both lowering the bed temperature and lowering the heat transfer coefficient increases the heat transfer surface and, thus, the bed depth, if the bed area is maintained constant. Any restrictions on bed depth would also have to be considered in the evaluation. Parametric curves have been prepared to enable this type of evaluation to be made.

The effect of boiler plant pressure drop and the steam-turbine condenser pressure on plant performance has been presented.[2] The boiler plant pressure drop has a small effect on plant capacity and heat rate: 1% increase in pressure drop results in a 0.1% increase in plant heat rate. An increase in the steam-turbine condenser pressure from 1-1/2 in. Hg. to 3 in. Hg. for a cooling tower results in a 2-1/2% increase in heat rate.

The potential performance of the plant was evaluated by assessing the effect of higher gas-turbine inlet temperatures and higher steam temperatures and pressures. The results indicate that plant efficiencies of ∿ 45% can be achieved with gas-turbine inlet temperatures greater than 2000°F with high-temperature blade material to minimize cooling requirements and with steam temperatures of 1200°F.

Advances in boiler plant subsystem concepts have not been considered in this analysis. Cost reductions may be achieved by using alternative concepts. The following components have been studied for potential savings:

- Particulate removal. The projected system utilizes four secondary collectors for final particle removal before the gas turbine. Alternative systems are being considered which may reduce the number and size of the units. Cost estimates indicate that reductions of $1 to $5/kw for the total particulate removal system may be possible.
- High-temperature gas piping. The high-temperature gas piping cost would be reduced if the particulate removal system were simplified by using fewer units per module. Additional savings might be realized if refractory lined pipe could be used between the secondary collectors and the gas turbine. The present design uses a high-alloy steel to assure protection of the gas turbine from additional particulates.
- Coal feeding system. The coal feeding system design is based on systems which have been built and operated. The design provides a separate coal feeding system for each fluidized bed in order to assure control of the coal feed rate to each bed. It may be possible, however, to reduce the number of coal feed systems from 16 to 4 if independent control of solids flow to each bed in a module can be achieved from a single pressurized injector. The potential cost reduction is estimated to be > $2/kw.

- Stack gas cooler design. Cost estimates were obtained for the stack gas coolers, but no attempt was made to optimize the design or consider nonconventional designs, such as those using fluidized beds. Preliminary conceptual evaluation indicates that the cost might be reduced from the present \$4.40/kw to ∿ \$3.40/kw.

2.7 CONCLUSIONS

- Base plant design is near optimal.
- The pressurized fluidized bed boiler power plant maintains more than a 5 to 10% energy cost advantage over a conventional plant with stack gas scrubbing.
  - Effect of a potential change in bed temperature, fluidizing velocity, heat transfer surface configuration, gas side heat transfer coefficient, boiler tube materials, bed depth limitations or pressure will result in
    - No significant change in plant operability or performance
    - < 2% increase in plant cost
    - < 0.2 mills/kWh increase in energy cost.
  - Effect of increasing the dust loading to three times the design value will increase the plant cost ∿ 4%, energy cost < 0.3 mills/kWh.
  - Effect of increasing excess air will result in
    - Increased turndown capability and performance
    - 3 to 6% increase in plant cost
    - No significant change in energy cost.
- Plant efficiencies of ∿ 45% may be achieved for gas-turbine inlet temperatures above 2000°F and steam temperatures of 1200°F.

3. LABORATORY SUPPORT STUDIES

3.1 INTRODUCTION

The evaluation of the fluidized bed combustion process for power generation carried out by Westinghouse concluded that pressurized combustion is the most promising technology from the point of view of

- Effective pollution abatement
- Efficient fuel utilization
- Economical power generation[1].

Among the key areas where further laboratory research and development were recommended were sulfur removal, stone regeneration, and sulfur recovery. In projecting possible sulfur removal, it was pointed out that, owing to a lack of experimental data, the effect of combustion pressure and of excess air on SO_2 reduction could not be predicted using the currently available models of the sulfur removal process.

Thus, the economical and ecological considerations which lead to the investigation of fluidized bed combustion as a clean source of electricity require the support of thermodynamic and kinetic data to translate the concept into a working system. Our extensive knowledge of the thermodynamics of the system contrasts sharply with the available kinetic data with which the system behavior can be predicted and interpreted.

The use of calcium carbonate (as limestone or dolomite) to prevent sulfur gas emissions in fluidized bed fuel processing depends upon a cycle of reactions, as shown in Figure 61. The conditions for the process options afforded by this reaction cycle are given in Table 17. Despite atmospheric pressure data obtained by Squires and his school on the reactions of hydrogen sulfide with dolomite, and various studies on sulfur dioxide sorption by limestones and dolomites, there is a

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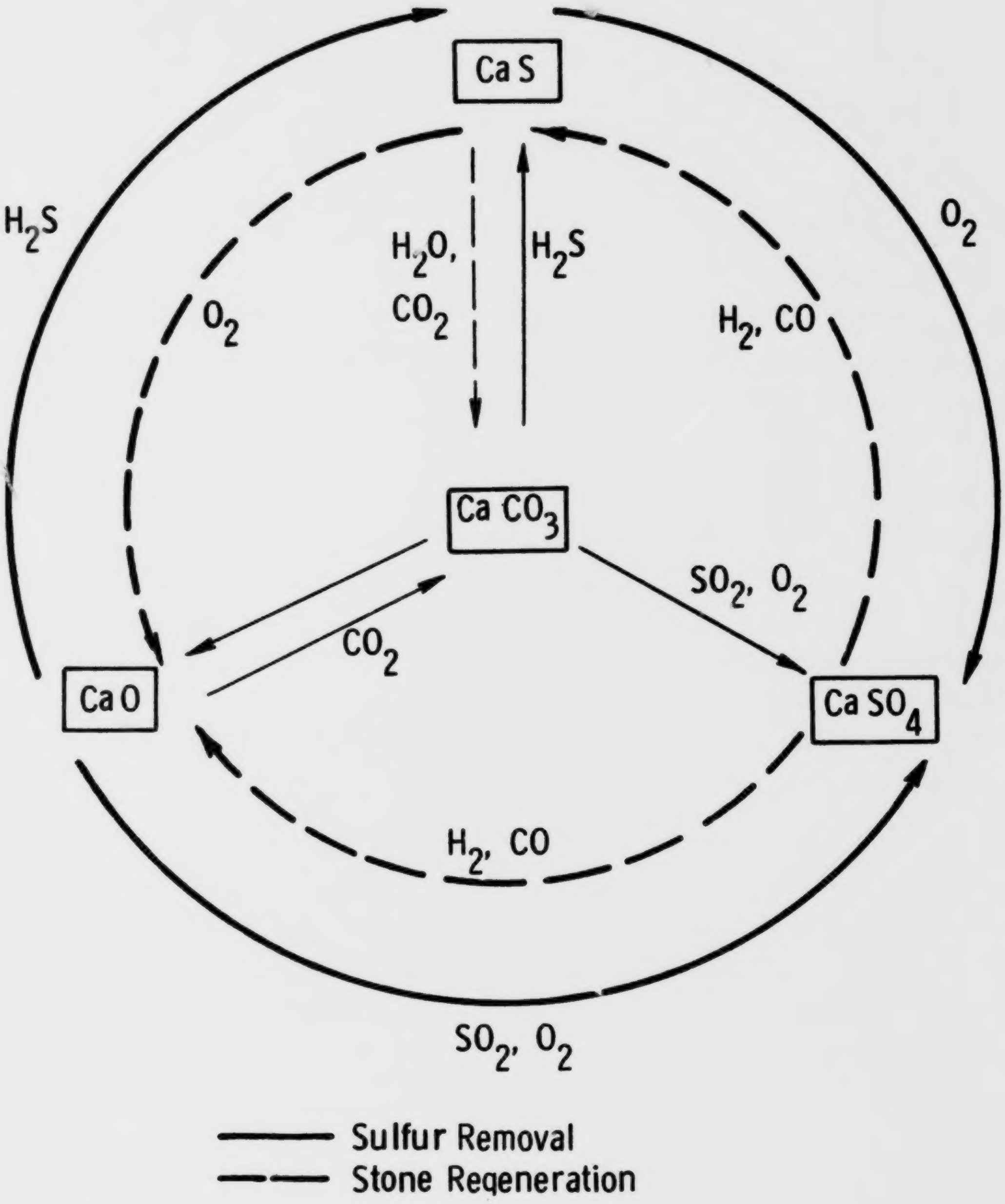
CaS
CaCO$_3$
CaO
CaSO$_4$
H$_2$S
O$_2$
O$_2$
H$_2$O, CO$_2$
H$_2$S
H$_2$, CO
SO$_2$, O$_2$
CO$_2$
H$_2$, CO
SO$_2$, O$_2$
Sulfur Removal
Stone Regeneration

Figure 61: The Calcium Carbonate/Sulfur Cycle Basic Reactions

TABLE 17

OPTIONS AFFORDED BY CALCIUM CARBONATE/SULFUR CYCLE BASIC REACTIONS

Reaction	Operating Conditions	Applicable Fuel Processing Option
Sulfur Removal		
1. $\begin{pmatrix}CaO\\CaCO_3\end{pmatrix} + SO_2 + 1/2\ O_2 \rightarrow CaSO_4 + \begin{pmatrix}-\\CO_2\end{pmatrix}$	1300°F < T < 1750°F P: 1 to 20 atm	combustion
2. $\begin{pmatrix}CaO\\CaCO_3\end{pmatrix} + H_2S \rightarrow CaS + H_2O + \begin{pmatrix}-\\CO_2\end{pmatrix}$	1300°F < T < 1700°F P: 1 to 20 atm	gasification
Stone Regeneration		
3. $CaSO_4 + \begin{pmatrix}H_2\\CO\end{pmatrix} \rightarrow CaO + SO_2 + \begin{pmatrix}H_2O\\CO_2\end{pmatrix}$	T > 2000°F	combustion (low SO_2 conc. at elevated pressure)
4. $CaSO_4 + \begin{pmatrix}4\ H_2\\4\ CO\end{pmatrix} \rightarrow CaS + \begin{pmatrix}4\ H_2O\\4\ CO_2\end{pmatrix}$, (followed by reaction 5, 6, 7 or 8)	T < 1600°F	combustion (at elevated pressure)
5. $CaS + H_2O + CO_2 \rightarrow CaCO_3 + H_2S$	T < 1500°F P > 5 atm	combustion/gasification (at elevated pressure)
6. $CaS + H_2O + CO_2 \rightarrow CaCO_3 + H_2S$	T ∿ 200°F atmospheric pressure	combustion/gasification (not recommended due to low temp. & water purification)
7. $CaS + 3\ CO_2 \rightarrow CaO + 3\ CO + SO_2$	T > 2000°F atmospheric pressure	combustion/gasification (not recommended due to low SO_2 conc., ∿ 2%)
8. $CaS + 3/2\ O_2 \rightarrow CaO + SO_2$	T > 1800°F	combustion/gasification (low SO_2 conc. at pressure)
9. $CaSO_4 + C + H_2O \rightarrow CaCO_3 + CO_2 + H_2S$		combustion (advanced concept)
10. $CaS + 2\ O_2 \rightarrow CaSO_4$	atmospheric pressure	all gasification

clear need for kinetic data with which the desulfurization of fuels in fluidized beds can be predicted and explained.

A thermogravimetric analysis system (TG) was modified so that it could be used to study the reactions of corrosive gases with solids at pressures above atmospheric. A program of studies on the sulfur/calcium carbonate cycle was then undertaken with the following objectives:

- To establish which reactions occur at proposed conditions
- To determine the reaction kinetics
- To determine the stone utilization achievable
- To study regeneration of the stone
- To probe reaction mechanisms
- To assess the influence of side reactions
- To recommend optimal operating conditions.

3.2 APPARATUS

The TG is an electrically recording balance which measures the changes in weight of a solid suspended from the balance arm in the electronically controlled furnace. The gaseous atmosphere surrounding the solid and the temperature of the furnace may be adjusted or altered as desired, within specifiable limits. Normal operating pressures are atmospheric or subatmospheric, and provision for the use of corrosive gases usually requires modification of commercial systems.

A number of investigators have reported work with a custom-built pressurized apparatus, as summarized in Table 18. No work appears to have been carried out with corrosive gases at pressures above atmospheric.

In our work, the DuPont thermal analysis system was modified for use on this project for three reasons:

- The successful use of corrosive gases in this apparatus by Squires, to study some of the reactions of interest

TABLE 18

PRESSURIZED TG SYSTEMS

Authors	Mechanism	Range	Studies	Kinetics
McKewan (1962)[7]	quartz spring	600 psi 1832°F	iron oxide reduction	Yes
Baker (1962)[6]	quartz spring		carbonate decomposition	(Thermodynamic Study)
Rabatin and Card (1959)[5]			$MnCO_3$ decomposition	
Johnson[8] (1968)	transducer	1000 psi	coal char gasification	Yes
Ho Bae (1972)[9]	transducer	1000 psi	oil volatilization	
Brown Penski (1971)[2] Callahan	DuPont	300 psig 662°F	evaporation; copper sulfate dehydration	
Wendland Williams (1972)[10] Simmons	DuPont	1000 psi	reduction of $K_3[Co(C_2O_4)_3]\cdot 3H_2O$	Yes
Squires (1973)[11]	no details available.			
Westinghouse (1972)	DuPont	600 psi 2190°F	$S/CaCO_3$ System	Yes

- The successful operation (under limited conditions) of a pressurized modification of the DuPont system by Brown and Penski
- The availability on-site of a DuPont system and technical expertise in its use.

3.2.1 Modification of the TG Apparatus

A DuPont 951 balance assembly and furnace (Figure 62) was mounted inside a pressure shell of stainless steel, as described by Brown et al.[2] Their design was modified to allow higher temperatures, a flowing gas in the reaction zone, and corrosive constituents in the gas.

High temperatures (2190°F) are possible because a web of copper cooling coils inside the pressure shell presses around the glass dome and metal block which houses the balance mechanism. A heat reflector (asbestos with an aluminum surface) further insulates the mechanism from the hot zone. External cooling coils wrapped around the pressure shell maintain its temperature at around 100°F even when the furnace is operating at high temperatures.

To protect the balance mechanism, a continuous flow of nitrogen sweeps out the chamber containing the balance mechanism. Reactant gases, which may contain H_2S or SO_2, are ducted into the quartz reactor containing the sample and after reaction unite with the nitrogen purge and are swept out of the reactor through a concentric outer chamber, as in the design of Ruth et al.[3,4] It is necessary to duct the gases out of the pressure shell, since the furnace windings are insufficiently protected by the insulating cement to prevent corrosion.

The gases are ducted from the modified quartz reactor through a stainless steel flexible tube sealed to the quartz with a rubber grommet. This seal is highly effective; and the grommet is replaced after

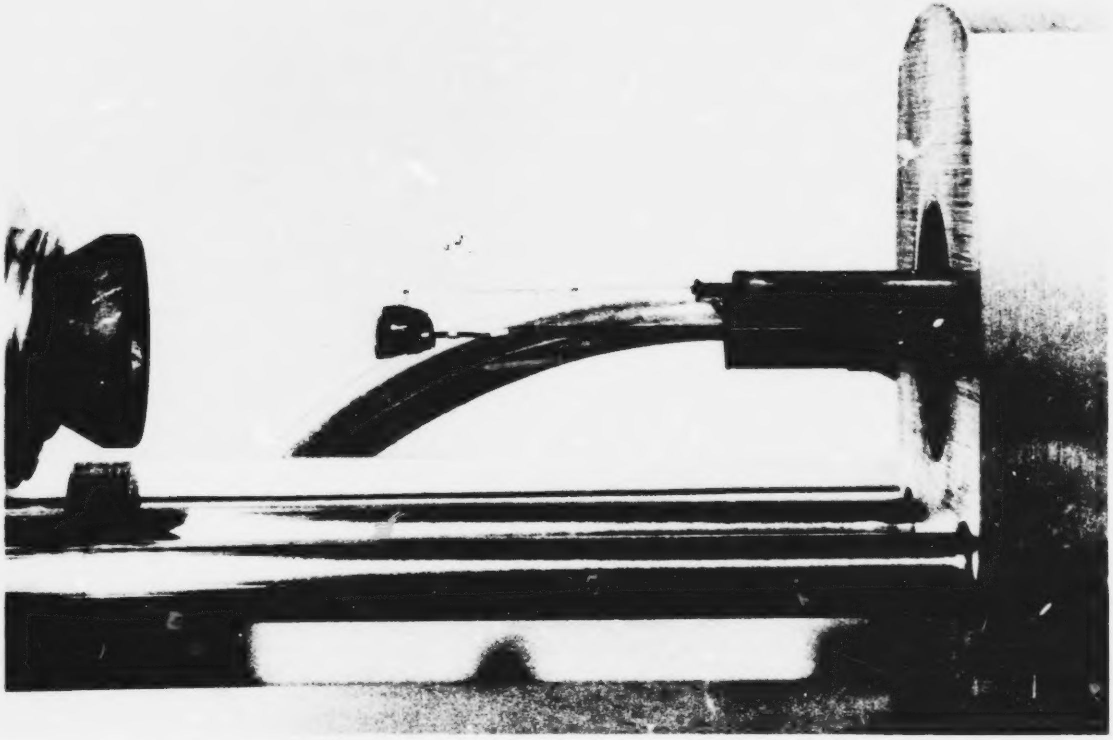

Figure 62: The Du Pont 950 Thermogravimetric Balance

each run. The exit gas line has provision for sampling, a trap for water, a zinc oxide trap for excess hydrogen sulfide, and a stainless steel filter, before it is depressurized through a pressure relief valve. The system is outlined in Figure 63.

Reaction gases are supplied from cylinders through precision stainless steel metering valves and flowmeters (five lines), and can by-pass the TG system while flow adjustments are under way. The quartz reactor tube abuts the end flange of the pressure shell and a thick rubber gasket makes the seal.

Steam was introduced where required by introducing a metered liquid stream through a fine steel capillary into a line heated by heating tapes. A thermocouple beside the steam generation point monitored the evenness of flow. A platinum wire heater mounted radially in the reactor where the reaction gas entered the end flange prevented condensation at this point,and the purge flow of nitrogen was sufficient to prevent the accumulation of liquid water at the cold end of the reactor.

The entire shell and TG was evacuable, and safety relief valves were mounted on the pressure shell and in the gas-flow exit line. Electrical connections for the furnace and the electronic controls entered the system via conax connectors.

A DuPont 990 thermal analyzer controlled the furnace and recorded the weight changes.

The system is shown in Figure 64.

3.2.2 Operation

Little modification in operating procedure is required. (Mounting the sample is tedious, but the operator's skill reduced the loading time to 15 minutes.) When the balance has been loaded, the end flange is bolted on and the system pressurized using the purge gas.

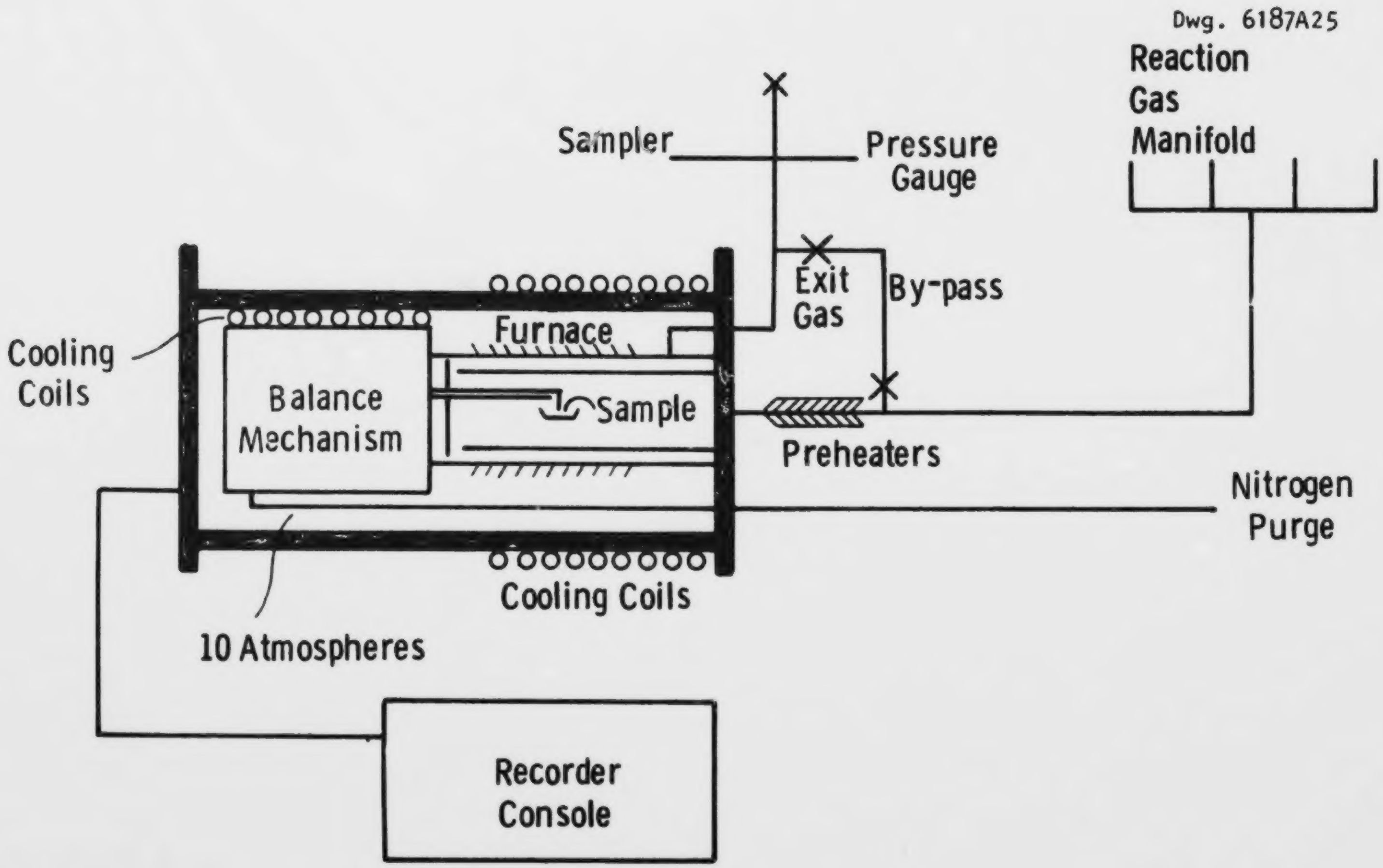
Dwg. 6187A25
Reaction Gas Manifold
Sampler
Pressure Gauge
Exit Gas
By-pass
Furnace
Cooling Coils
Balance Mechanism
Sample
Preheaters
Nitrogen Purge
Cooling Coils
10 Atmospheres
Recorder Console

Figure 63: Diaqram of the TG System

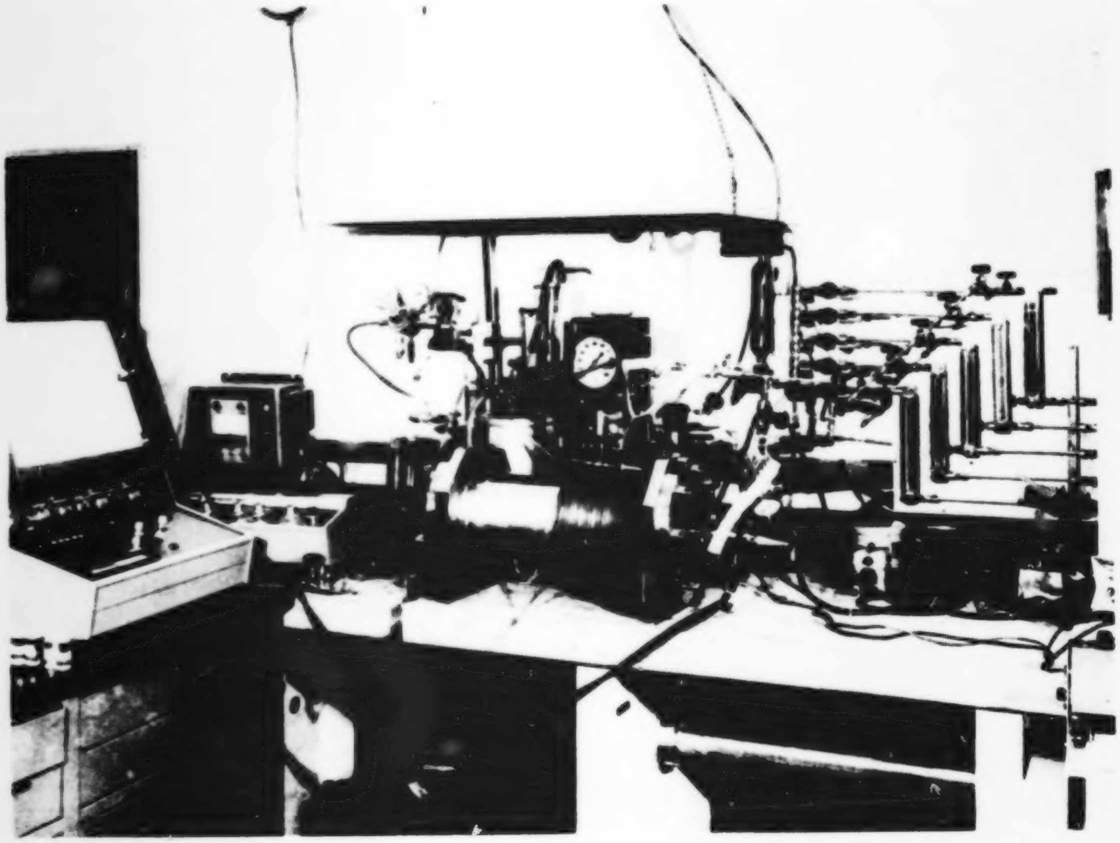

Figure 64: The Pressurized TG Apparatus

Buoyancy effects are not a problem at temperatures greater than 1100°F; changes in the reaction gas stream inevitably lead to minor buoyancy effects which can be determined experimentally and compensated for in data interpretation. Temperature stability is not as good as during atmospheric operation: the Cr/Al thermocouple is of necessity sheathed in a quartz capillary when corrosive gases are in use, and the flow of reactant gas places a duty on the furnace which leads to temperature inhomogeneities in the furnace. Changes in the reactant gas composition result in changes in the isothermal temperature recorded by the thermocouple, lending a degree of imprecision to the kinetic data.

Because of the heavier duty imposed on the furnace, resulting in less precise temperature control, and the quartz shield on the thermocouple, Ⓦ checked the accuracy of the temperature recording. A sample of specpure iron was heated in nitrogen at 5 K/min. The observed Curie point was 1424°F, in satisfactory agreement with the literature value, 1418°F.

3.2.3 System Performance

The performance of the apparatus may be assessed in terms of the quality of the data produced and the reliability of the instrument.

The data obtained during atmospheric pressure operation proved to be repeatable. An example may be cited where a cycle of reactions was performed on an iron oxide substrate. The sample was reduced, sulfided, oxidized, and reduced for seven cycles. Each cycle required that the gas flow be reset. The mean deviation of the rate at 5% utilization was 3.6%, from the second to the seventh cycle. It is more difficult to assess the repeatability of the experiments involving small samples of impure limestone and dolomite. Buoyancy effects were not serious at temperatures above 1100°F; the precision of weight measurements is best demonstrated by the agreement obtained in weight changes. For TG124 where the reactions

$$CaS + H_2O + CO_2 \rightarrow CaCO_3 + H_2S \qquad (1)$$

$$CaCO_3 \longrightarrow CaO + CO_2 \qquad (2)$$

occurred, the weight changes observed corresponded to:

(1) 1.45×10^{-4} moles carbonate formed
(2) 1.43×10^{-4} moles carbonate decomposed.

Two forms of breakdown restricted the number of experimental runs. Corrosion of the balance mechanism requiring repair by DuPont occurred twice. It is thought that corrosion was accelerated when the pressure relief valve became blocked. When gas flow ceased in the exit line, accumulation of hydrogen sulfide in the system overcame the protection afforded by the purge gas. The material which accumulated in the seat of the pressure relief valve was, apparently, solid sulfur. A fine powder of sulfur accumulated in the valve despite the presence of a stainless steel filter in the exit line. No sulfur accumulation was noted elsewhere in the exit line. In an effort to overcome this problem, a zinc oxide trap heated to 572°F was inserted in the exit line, before the pressure relief valve. This has prevented any sulfur deposition in the valve, when gases containing hydrogen sulfide are used.

Electrical faults developed on two occasions in the first eight months of operation of the 990 console, necessitating replacement of two circuit boards and a relay. In the second breakdown, additional delay was caused by a faulty replacement circuit board.

3.3 PROCESS VARIABLES

The review of experimental results in the evaluation of the fluidized-bed combustion process[1] listed thirteen variables which influence the reduction of sulfur oxide emission. Since some of those variables are in part complex functions of more fundamental physical variables, they can be associated with factors which influence the gas-solid kinetics.

Fluidized Bed	Kinetic Study
Combustion pressure	Pressure
Bed depth	SO_2 concentration
Ca/S mole ratio	
Sulfur level of coal	
Bed temperature	Temperature
Fluidizing velocity	Controlling regime (external mass transfer, internal mass transfer, chemical reaction
Factors affecting gas-solids mixing	
Stone type and source	Stone type
Stone size	Particle size
Excess air	O_2 concentration
Coal type	
Location of coal feed	Oscillation from oxidizing to reducing atmosphere

The variables on the right have effects which can be studied in the TG apparatus. The major consideration in this study is the effect of pressure in combination with the other variables.

Some factors are of particular concern. The operating temperature of the pressurized fluidized bed design is 1750°F (chosen in consideration of desirable turbine inlet temperature and ash softening characteristics). However, the retention of sulfur in lime-containing fluidized beds has shown a maximum at lower temperatures in some fluidized bed combustion experiments.[14,15,16] The variation in the rate of reaction of SO_2 and CaO with temperature is , therefore, of great interest. The rate of reaction and the degree of stone utilization achievable are also of primary importance. An increase in the rate due to pressurized operation may lead to better sulfur retention: increased utilization of the actual CaO content affects the criteria for comparing once-through systems with regenerable sorbent processes.

3.4 PREVIOUS LABORATORY STUDIES

The results of several laboratory studies are summarized here.[14-20]

- Sulfation is first order with respect to SO_2.
- Sulfation is zero order with respect to O_2.
- The rate of reaction increases with temperature. Various values of activation energy have been derived, but there is no unambiguous determination of activation energy.
- Reactivity depends upon the pore volume formed during calcination.
  - Small pores give a high initial reaction rate but, if the particles are large, a low overall capacity results.
  - Large pores give lower rates but increased capacities.
  - For small particles, capacity is determined by the pore volume available for product accumulation.
  - Other properties which influence the pore structure formed on calcination may serve as general indices to the capacity, i.e., temperature and conditions of calcination and sodium content of the stone.
  - The distribution of sulfur through a sulfated stone depends on the stone: a coarse limestone shows total permeation of the stone by sulfate, while iceland spar forms a rim of sulfate.

3.5 SUMMARY OF THE RESULTS

3.5.1 Pressure

Sulfation at 10 atmospheres was not significantly faster than sulfation at 1 atmosphere. However, the utilization achieved before the rate fell to one-fiftieth of its initial value was 84% at pressure as compared to 42% at 1 atmosphere, when the stone was calcined above 1560°F under an atmosphere of CO_2.

3.5.2 Sulfur Dioxide Concentration

The sulfation was found to be first order with respect to sulfur dioxide concentration. The increase in partial pressure of sulfur dioxide in passing from 1 to 10 atmospheres at fixed gas composition does not result in a ten-fold increase in observed rate. Changes in the concentration of sulfur dioxide at a given pressure produce a proportional change in rate.

3.5.3 Temperature

A slight increase in rate is observable with temperature in the range 1350 to 1700°F for dolomite 1337 (Figure 65); for Tymochtee dolomite initial rates (up to 25% utilization) are not distinguishable in the range 1600 to 1850°F, but utilization is ultimately lower at the higher temperature.

3.5.4 Agreement with Previously Reported Data

The kinetics observed on the TG apparatus were sensitive to sample size: for small samples (10 mg raw dolomite) at atmospheric pressure, the rates of reaction were in good agreement with those found by Borgwardt and by the NCB investigators.

3.5.5 Stone Type

Both dolomite 1337 (high purity) and Tymochtee dolomite (10% impurities) proved capable of yielding high utilization.

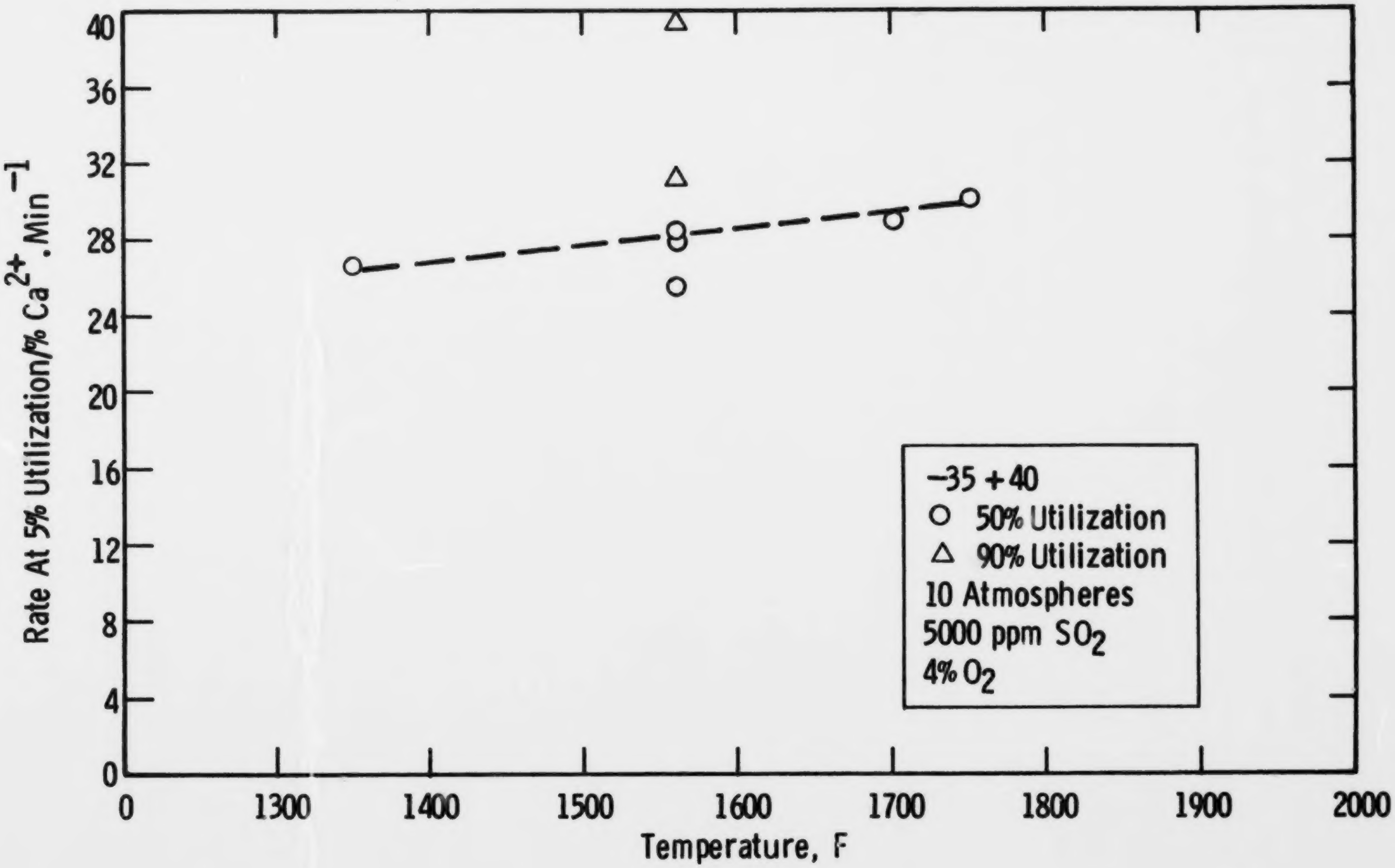

Figure 65: Temperature Effect on Rate for 1337 Dolomite Sulfation

3.5.6 Particle Size

Stone reactivity decreased as particle size increased, but the rate of reaction of 4000 μ diameter particles was faster than shell diffusion would permit, indicating a contribution to reaction from internal surfaces.

3.5.7 Oxygen Concentration

The reaction was zero-order in oxygen concentration in the range 2 to 4%. Increasing the oxygen concentration to 11% gave no increase in rate.

3.5.8 Stone Pretreatment

The result of stone pretreatment was crucial in its effect on CaO utilization.

When calcination of the stone was suppressed with carbon dioxide until temperatures of ∿ 1560°F were achieved, subsequent calcination yielded an extremely active stone, as shown in Figure 66.

Attempts to vary the calcination by slow calcination at lower temperatures led to minor variations in the utilization.

The effect of calcination under CO_2 for Tymochtee dolomite is shown in Figure 67. The active and inactive stone yielded 40% CaO utilization within twenty minutes, although the active stone was exposed to one-seventh of the SO_2 concentration used in the inactive stone experiment.

3.6 THE EXPERIMENTAL PROGRAM

3.6.1 Sulfation of Dolomite

In the first phase of the experimental program, the sulfation of two dolomites at 10 atmospheres pressure was studied. Analysis of the rate data led us to undertake experiments at atmospheric pressure to establish the relation between our results and those obtained by other workers.

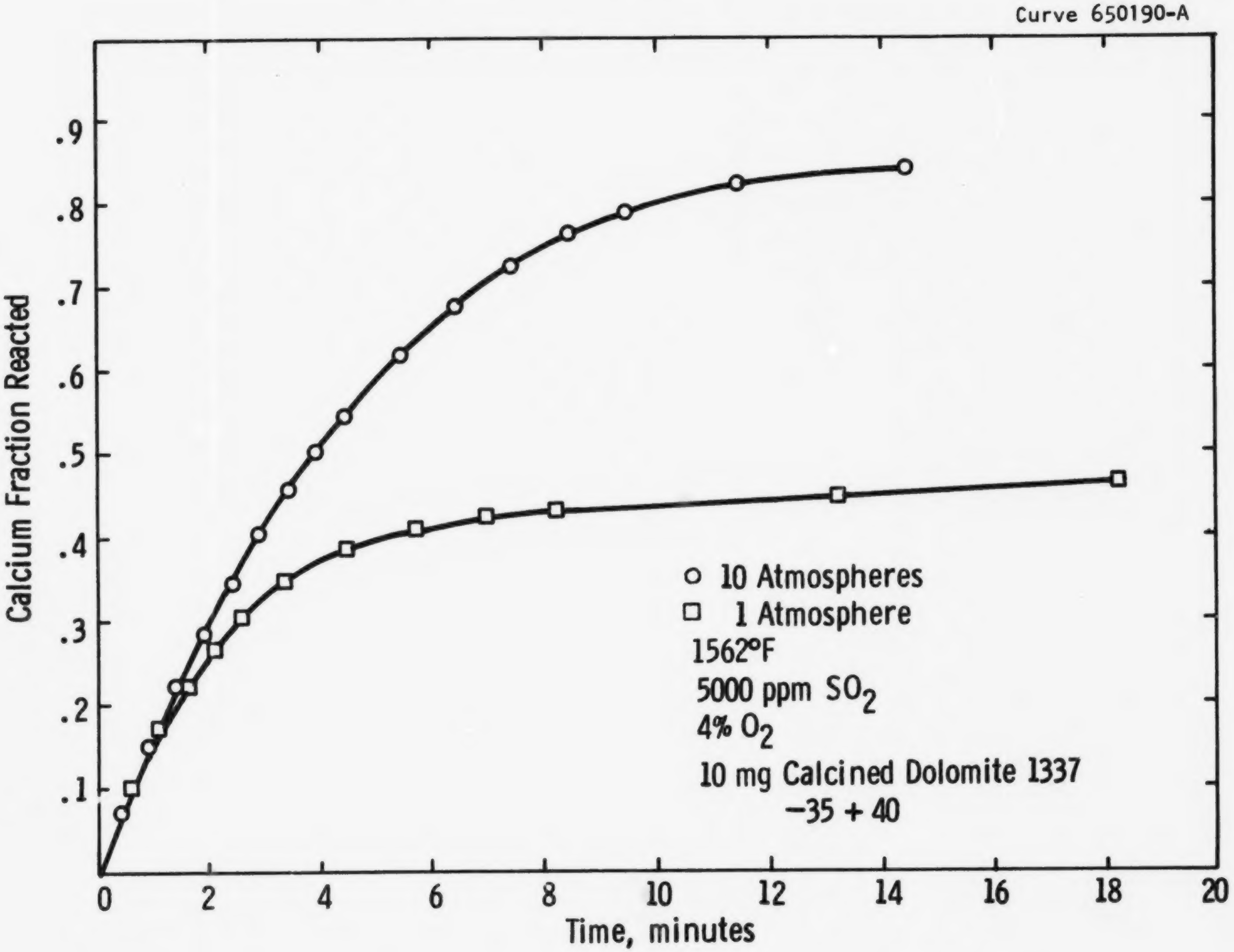

Figure 66: Sulfation of Calcined Dolomite

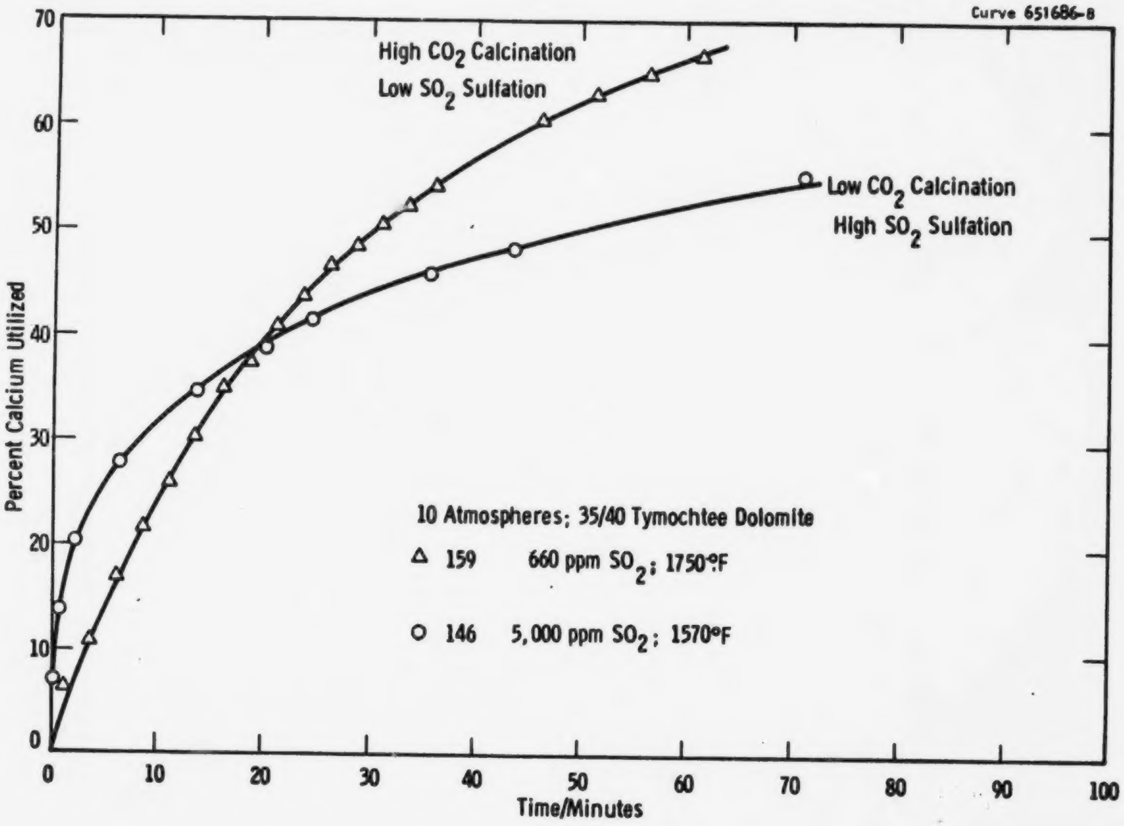

Figure 67: The Effect of Calcination Conditions on Sulfation

Samples of dolomite weighing 30 to 100 mg were calcined and sulfated (30/40 mesh being used). The experimental details for experiments TG86-TG98 are shown in Tables 24 and 25.

Calcined Tymochtee dolomite reacted at a rate dependent on calcium utilization. At 1562°F, in 4% O_2 and 1250 ppm SO_2, 25% utilization was reached within eleven minutes, but exposure to reactant gas for one hour was needed to achieve 50% utilization.

The course of reaction followed a rate-law of the form identified by previous investigators:[18]

$$\frac{d\alpha}{dt} = \frac{a}{t_o} e^{-\alpha/a}$$

as shown in Figure 68. Basically, this rate-law describes a transition between an initial fast rate and a much slower rate which eventually controls reaction.

Calcined dolomite 1337 exhibited the same behavior on sulfation. However, the initial fast rate was one-third of that reported by Borgwardt for the same dolomite. The rate of reaction for half-calcined stone was the same as that for fully-calcined stone, and increasing the gas-flow rate from two to four liters per minute had no effect on the rate.

It was concluded that the geometry of large samples (30 to 100 mg) might prevent an adequate supply of reactant gas from reaching the sample. Therefore, atmospheric pressure tests were carried out to permit more direct comparison with previous work reported by other investigators.

3.6.2 Atmospheric Pressure Tests (TG102-108)

Atmospheric pressure sulfation of calcined 1337 dolomite showed that

- Sample size influenced the rate of reaction; large samples exhibited slower rates than small samples (Figure 69).

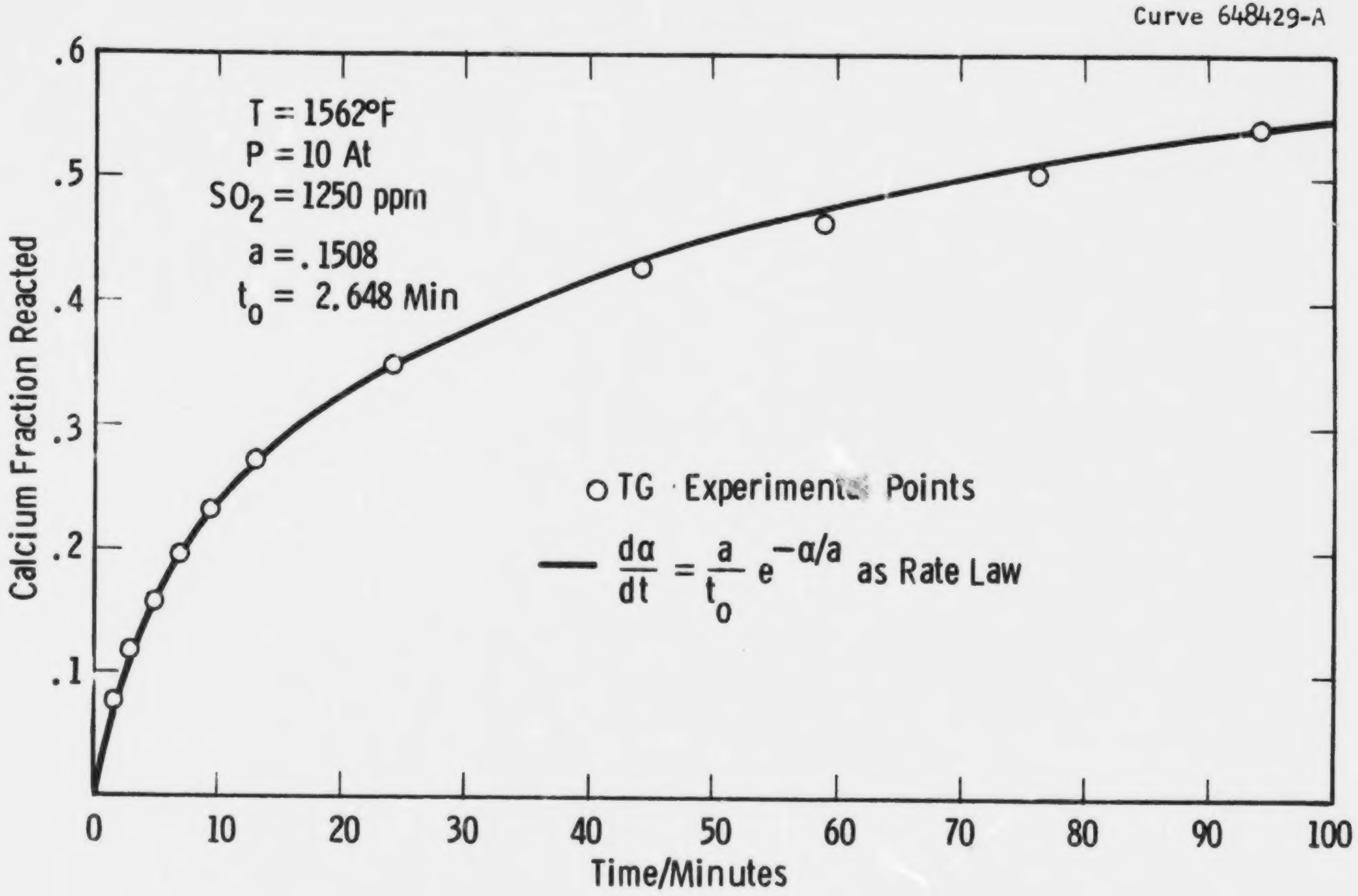

Figure 68: Sulfation of Calcined Tymochtee Dolomite TG 90B

Curve 648706-A

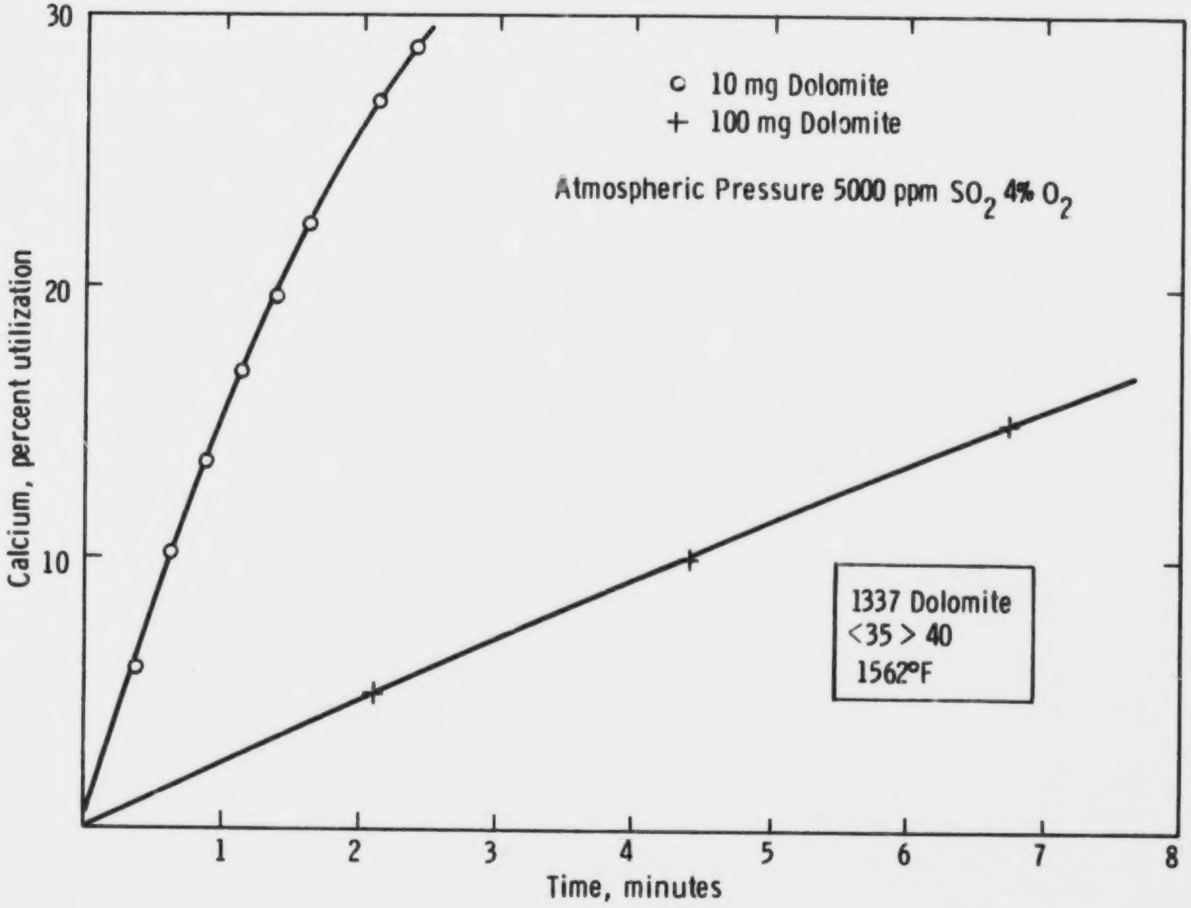

Figure 69: Effect of Sample Size on Sulfation of Calcined Dolomite 1337 in the TGA

- The flow rate of reaction gas had only a minor effect on the initial rate of reaction (Table 19).
- The rate of reaction observed at atmospheric pressure was close to that estimated from the graphs published by Borgwardt, when the sample size was kept to 10 mg (Table 19).
- Although the smaller sample size showed a faster rate of reaction, termination of the rapid phase of reaction was more abrupt. Thus after thirty-four minutes, the same degree of sulfation was achieved in the case of both samples, and the larger sample proceeded from there to react at a faster rate (Figure 70).
- A small sample (10 mg) was sulfated at 10 atmospheres pressure, and it exhibited a high initial rate of sulfation, marginally faster than the rate at atmospheric pressure under corresponding conditions. However, it did not undergo an abrupt transition to a slower rate of reaction at 43% sulfation, and reached over 90% of theoretical uptake (Figure 71).

The principal discoveries of the preceding experiments can be summarized as:

- At atmospheric pressure, the rate of reaction on the TG is similar to that observed by other workers.
- Large samples hinder the determination of maximum reaction rate.
- At pressure (10 atmospheres), the rate of reaction does not increase greatly. Utilization of the calcium may be far greater at pressure (> 90% conversion of CaO to $CaSO_4$, in dolomite), than at atmospheric pressure (∿ 50% conversion of CaO to $CaSO_4$, in dolomite).

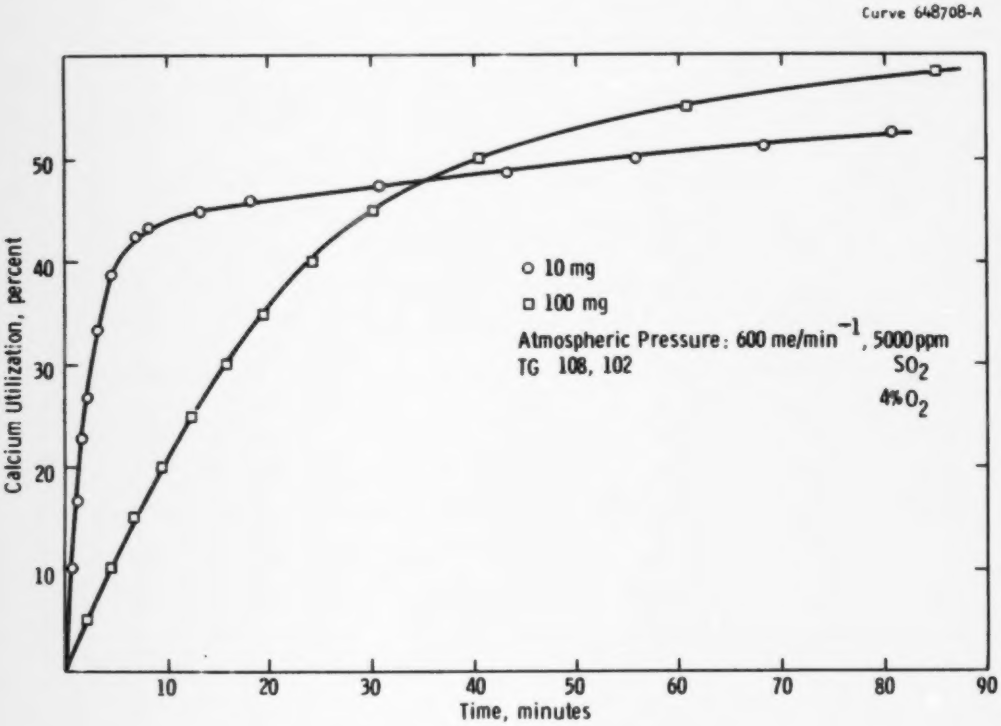

Figure 70: Sulfation of Calcined Dolomite 1337

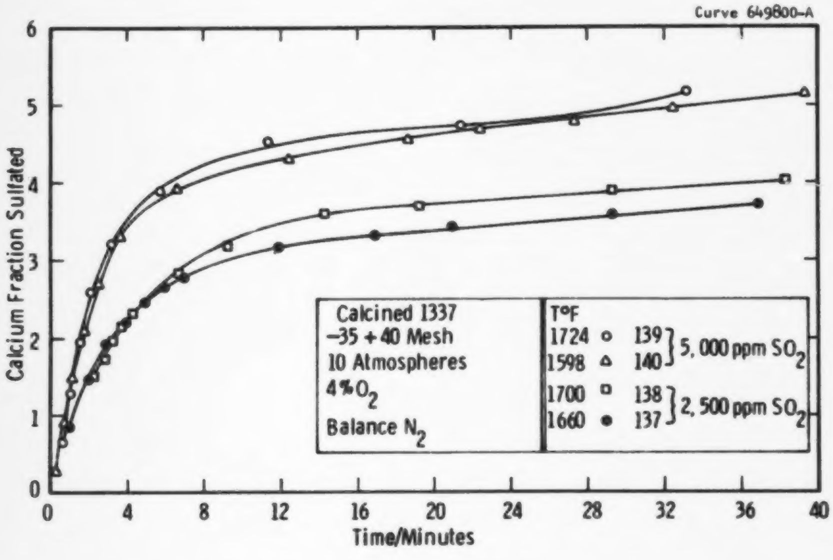

Figure 71: Sulfation of Calcined Dolomite 1337

TABLE 19

RATE OF SULFATION OF CALCINED DOLOMITE 1337, AT 5% CALCIUM UTILIZATION, 1562°F

TG Expt. No.	Pressure (atm)	Reaction Gas Flow Rate ml. min^{-1}	Reaction Rate $10^5 \times \frac{\text{mole SO}_3}{\text{(g calcined dolomite)(sec)}}$	Comment
107	1	100	1.77	10 mg sample
106	1	200	1.70	10 mg sample
105	1	400	1.81	10 mg sample
108	1	600	2.40	10 mg sample
103	1	400	0.46	large sample (100 mg)
110	10	2000	2.72	10 mg sample
---	1	---	2.1	from Borgwardt's graphs at 1600°F

- Since the calcination history of large samples is different from that of small samples, the mode of calcination may govern the utilization achievable.

3.6.3 Dolomite 1337 at 10 Atmospheres (TG137-150)

In these experiments, the following parameters were varied: SO_2 concentration; temperature; calcination treatment.

Temperature had almost no effect on the rate of reaction in the range 1350 to 1150°F as shown in Figure 65.

The change from fast reaction to slow reaction occurred at 30% utilization at 2500 ppm SO_2 and at 40% utilization at 5000 ppm SO_2, as shown in Figure 71. The utilization in these experiments was similar to that observed earlier in the atmospheric pressure experiments.

The SO_2 concentration in the slow phase was varied from 4000 to 1000 ppm, and held at a given concentration until the rate was established. The rate was first-order in SO_2 concentration (Figure 72).

The calcination treatment prior to sulfation was varied from rapid calcination at 1472°F in nitrogen to slow calcination by heating from 1112°F at 2 K/min.

Slow calcination, by heating at 2 K/min, combined with reaction at 1750°F decreased the utilization observed in the fast reaction period 30%. Lowering the temperature of sulfation to 1680°F failed to restore utilization after slow calcination (a lower SO_2 concentration was used).

Finally, a calcination treatment in which the magnesium fraction was permitted to calcine but the calcium carbonate was held under 1 atmosphere of CO_2 was tried. Calcination of the carbonate occurred rapidly at 1690°F. Sulfation of the stone so produced gave high utilization in the fast rate phase and was comparable to the result observed in TG110 (Figure 73).

Thus, minor differences in the utilization achieved during the fast period of reaction result from SO_2 concentration and temperature

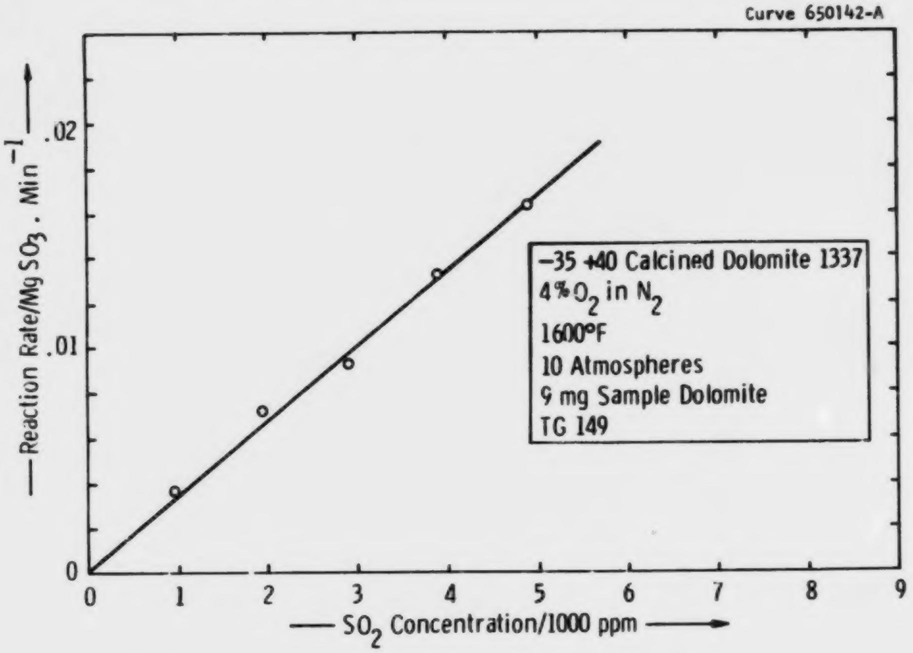

Figure 72: Effect of SO_2 Concentration on the "Slow" Sulfation Regime

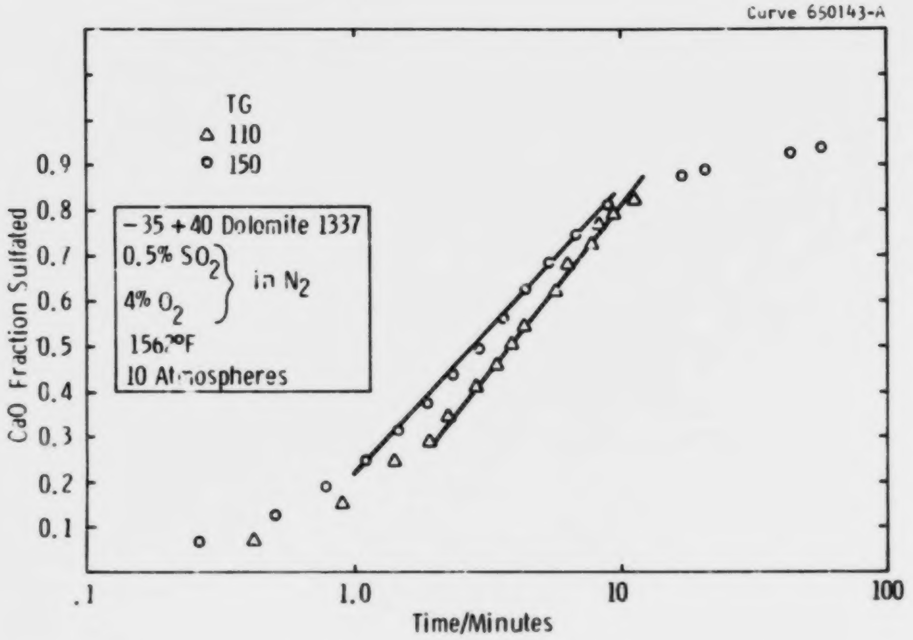

Figure 73: High CaO Utilization in the Sulfation of Calcined Dolomite

effects. The effect of calcination can produce differences of the order of 10% utilization for variations in the calcination heating rate; or differences of 40% when calcination is carried out at 1652°F under CO_2.

3.6.4 Sulfation of Tymochtee Dolomite at 10 Atmospheres

The remaining experiments were carried out using Tymochtee dolomite. The calcination treatment which had been successful in leading to high utilization was used. A larger particle size (16/18 mesh; 1000 μ diameter) was used, and in three experiments the temperature of sulfation was varied from 1600°F to 1850°F. Three features of the course of reaction was notable (Figure 74).

- High utilization was achieved (> 60% in 20 minutes).
- The rates of reaction at the three temperatures were not distinguishable in the early part of the reaction.
- After 20% utilization, the rate of reaction decreased with temperature, leading to a 10% difference in utilization between the 1600°F and the 1850°F sample. This difference may be noted as an increase in the time required to reach a given level of utilization and is shown in Table 20.

3.6.5 Effect of Low SO_2 Levels

In further experiments, the effect of high oxygen concentrations and low SO_2 levels on the reaction were investigated. Tymochtee dolomite was used as substrate, and calcination was carried out at 1562°F to attempt to get high utilization.

Three experiments at 660 ppm SO_2 and 1750°F were carried out on three different particle sizes, and one experiment at 90 ppm SO_2 was performed on one particle size (Figures 75 and 76).

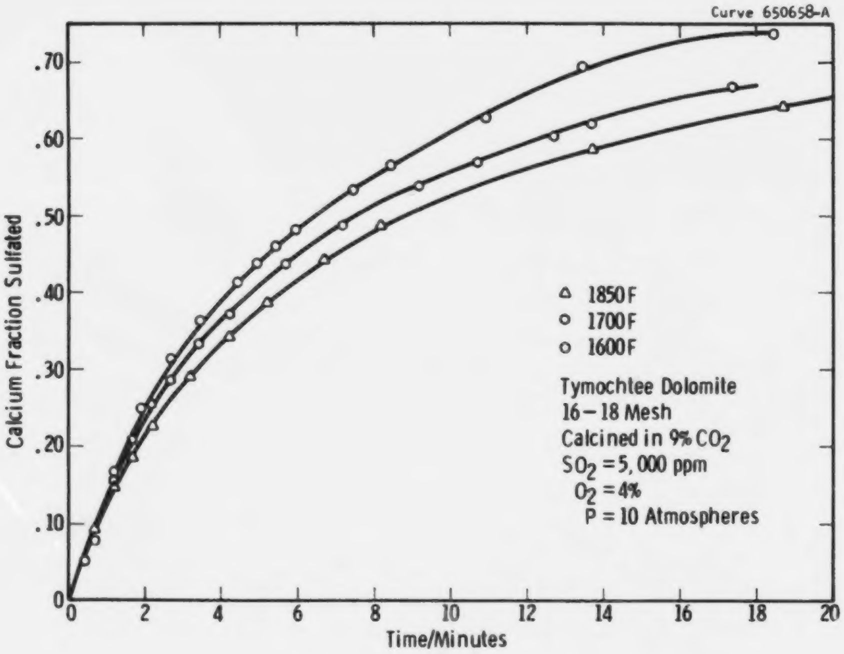

Figure 74: Tymochtee Dolomite Sulfation under Pressure: The Effect of Temperature

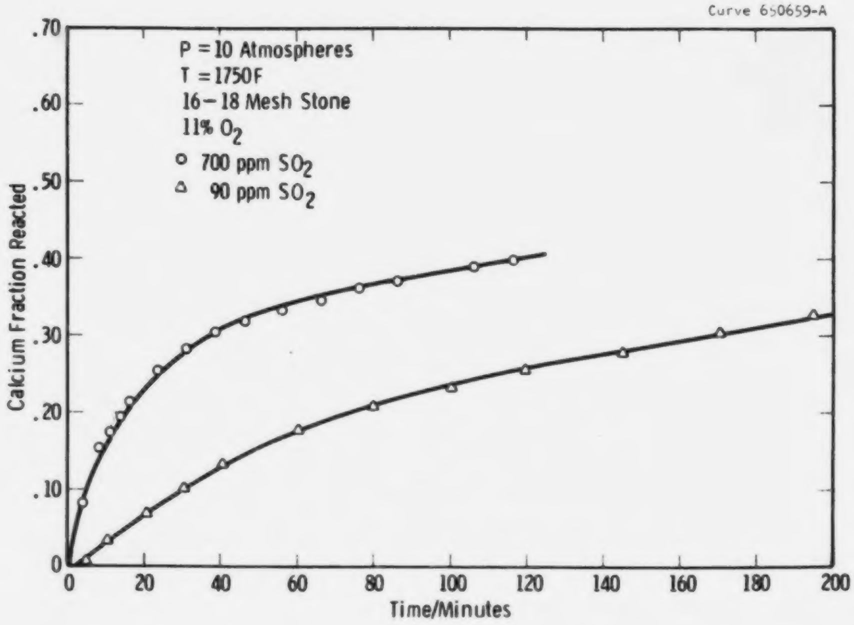

Figure 75: Sulfation of Calcined Tymochtee Dolomite at Low SO_2 Concentrations

TABLE 20

THE EFFECT OF TEMPERATURE ON THE TIME REQUIRED TO ACHIEVE A GIVEN LEVEL OF CALCIUM UTILIZATION

Utilization % Ca	Time at 1600°F	Time at 1850°F	% Increase in Reaction Time
25	1.99	2.5	25
35	3.4	4.4	29
45	5.2	7.0	35
55	7.95	11.3	42
65	11.55	19.45	68

Curve 650660-A

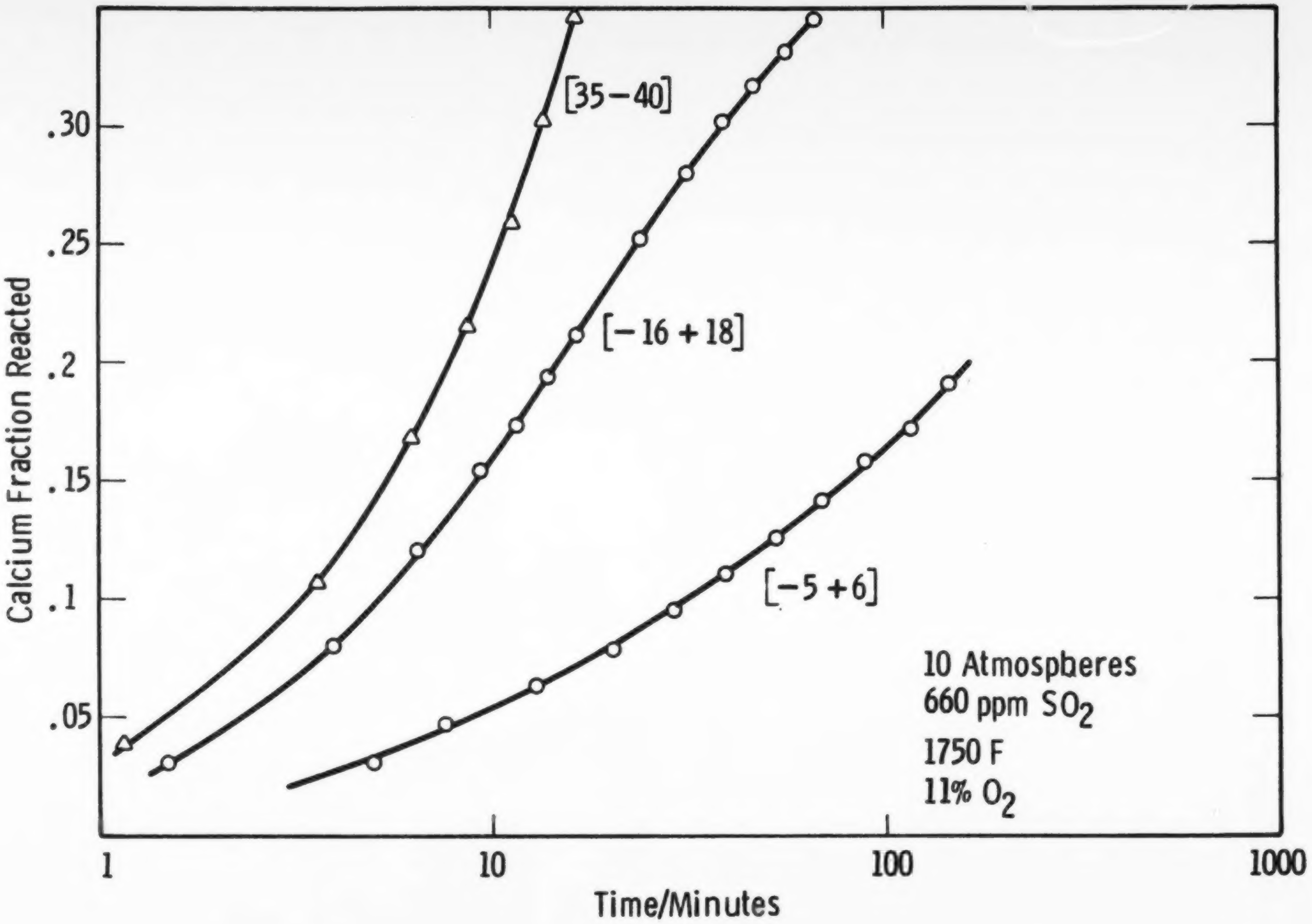

Figure 76: Sulfation of Calcined Tymochtee Dolomite

The experiment at the lowest SO_2 concentration showed that the course of reaction was in reasonable accord with expectation, on the basis of the reaction behavior at 5000 ppm. Table 21 shows the rate of reaction relative to unit rate at 5000 ppm, at five values of calcium utilization. The 660 ppm results were anomalous, in that the initial rate was in agreement with first order predictions, but as reaction proceeded, the rate fell off more sharply, as if high utilization was not to be achieved. Since the calcination conditions may not have yielded an active structure, the utilization achieved (Figure 75) is pessimistic for the 1000 μd stone.

Comparison between the 500 μ stones sulfated at two different SO_2 concentrations clearly illustrate that high utilization can be achieved with the Tymochtee stone.

The results obtained in these experiments are relevant to the design of the fluidized bed adiabatic combustor combined cycle power plant. In this concept, excess air in the fluid bed ($\sim$ 300%) effectively provides the heat transfer surface. Desulfurization in the fluidized bed is affected both by higher oxygen concentrations (13%) and by lower sulfur dioxide concentrations ($\sim$ 1000 ppm or less).

The results obtained on the TG show that no increase in rate can be expected as a result of the increased partial pressure of oxygen. This means that the step in the sulfation sequence which involves oxygen is not rate-controlling under these conditions. The dilution of sulfur dioxide reduces the rate of reaction, but clearly a sufficiently long residence time of stone in the bed will lead to efficient calcium utilization, and satisfactory sulfur removal will depend on the bed depth chosen. In a later section, the TG data reported here is used to program Koppel's model for fluidized bed conditions, and an estimate of the bed height required is obtained.

It is interesting to note that even at the 90 ppm SO_2 level, reaction proceeds at a nonnegligible rate, and 30% calcium utilization is noted after three hours' residence of a 1000 μ diameter fraction in such a stream. This is roughly equivalent to a 10% reduction in SO_2 concentration per centimeter of bed height at the 90 ppm level, at 4 ft/sec through the bed.

TABLE 21

THE RATE OF SULFATION OF TYMOCHTEE DOLOMITE AT LOW SO_2 CONCENTRATIONS, 1750°F

% Ca Utilization	5000 ppm	660 ppm	90 ppm
5	1	.132	.025
10	1	.168	.026
20	1	.072	.017
30	1	.048	.016
35	1	.030	.024
SO_2 ratio	1	.132	.018

3.7 THEORY

The effects of pressure on gas/solid reaction kinetics derive from two fundamental and competing phenomena. Increasing the reactant gas pressure increases the number of reactant molecules striking the solid surface in unit time, but it decreases the diffusion of products away from the reaction site. In addition, where reaction is part of a chain that includes adsorption or desorption as a rate-influencing step, saturation of the available sites can place a limit on the influence of pressure on the reaction rate.

Studies on the reduction of iron oxide have demonstrated the usefulness and limitations of pressurization. Thus, doubling the hydrogen pressure was found to lessen the time required to reduce a given sample of FeO, but it did not shorten the time for reduction to 50% of its former value.[12] McKewan[13] found that an increase in pressure from 1 to 10 atmospheres increased the rate of reduction of magnetite with hydrogen by a factor of 1.88.

These results may be interpreted as arising from convective mass transfer control at the gas/solid interface or as an adsorption-controlled phenomenon, limited by the number of available surface sites. Distinction between the two mechanisms may reflect the interests of the chemist or engineer studying the reaction rather than an intrinsically different behavior at the molecular level.

3.8 KINETICS OF DOLOMITE SULFATION

The kinetics of dolomite sulfation may be described by the rates of individual steps as follows:

1) Arrival of SO_2 at the particle surface
2) Transport of SO_2 into the dolomite
3) Transport of SO_2 across a product layer of $CaSO_4$ on the reactant crystallites of CaO
4) Chemical reaction between SO_2, CaO and O_2 (itself a multistep process).

The rate of the fourth step represents the maximum achievable rate; to attain it, the rates of the other three processes must be maximized. Step 1 will depend on the gas/solid mixing or on the gas velocity over the particle. Steps 2 and 3 depend on the physical composition of the stone. Thus, if large pores are available to transport the gas into the stone, then step 2 will not limit the rate of reaction. Similarly, large pores will facilitate the expansion required to accommodate the sulfate ion. Step 3 will begin to control the rate at some stage as reaction proceeds, but maximizing the surface area of CaO exposed to the gas (small CaO particles) will ensure a large extent of reaction before solid product layers inhibit the rate. The aim of this investigation is, in effect, to find the conditions whereby the rates of steps 2 and 3 are maximized. Since the structural features of the stone which are important in steps 2 and 3 are formed during calcination, it is clearly important to know what happens during calcination.

Dolomite calcines to either the half-burned state $MgO \cdot CaCO_3$, or the fully-burned state $MgO \cdot CaO$.

In the half-burned state, the external morphology remains the same, but the surface is covered by minute 400 Å cubic crystallites of MgO.

The start of calcite decomposition is independent of the presence of magnesia (but only at the lowest pressures does the decomposition of the calcite approach completion) since MgO covering the calcite may sinter and grow in CO_2 at the higher temperatures. In the half-burned state, the MgO crystallites are small - ∿ 150 Å, while the calcite crystals are about 400 Å. As the temperature is increased, their size passes through a minimum near 1478°F and back to 400 Å at 1650°F. However, in the fully burned state, misoriented microcrystals of both oxides form directly, their dimensions rise continuously from 100 Å at 1112°F to 1400 Å above 1830°F, and a reduction in pore volume also occurs over the same range, the smaller capillaries ∿ 80 Å being eliminated first.[21]

Sulfation will, therefore, exhibit kinetic behavior characteristic of the stone sorbent structure laid down during calcination. The effect of carrying out the reaction at 10 atmospheres pressure will reflect the effect of pressure on steps 1 to 4. Thus the effect of pressure on the transfer of SO_2 from the gas phase to the solid will increase it by a factor of two, given the experimental conditions used here. The transport of SO_2 into the dolomite will be largely unaffected by pressure, since gas concentration is proportional to pressure and the diffusion coefficient is inversely proportional to pressure. The effect on transport of SO_2 across a layer of calcium sulfate depends on the actual mechanism. The effect of pressure on the chemical reaction step should be to increase the rate ten-fold. Thus the combination of processes may result in acceleration of the rate of 10 atmospheres from one to ten times its atmospheric pressure value.

The experimental rate observed in these experiments is slightly higher than those observed at atmospheric pressure; the overall reaction is extremely similar (Figure 71), demonstrating that the reaction is controlled by a process virtually independent of pressure.

3.9 QUALITATIVE COMPARISON WITH FLUIDIZED BED RESULTS

The atmospheric pressure results reported here agree with data reported by other laboratory investigators. It is, therefore, instructive to compare this laboratory data with experiments on real or simulated sulfur retention in fluidized bed combustors.

In our atmospheric pressure experiments on 500 μ diameter particles of dolomite 1337, the rate of reaction typically falls to 1/50th of its initial value at 42% utilization. Borgwardt's experiments on similarly sized particles of the same dolomite showed that 55% utilization occurred for the same decrease in rate. In the NCB experiments, similarly sized particles gave 52% utilization within 1/4 hour, and 54% utilization in 4 hours. Borgwardt's experiments on 100 μ diameter particles showed that about 84% utilization occurred before the rate fell to 1/50th of its former value.

These results may be compared with the utilization achieved by Esso (R & E), Argonne National Laboratory, and the National Coal Board (NCB) with dolomite 1337 in fluidized beds.

Argonne's test in a 6-in. fluidized bed combustor (BC 10), using 540 μ dolomite 1337 showed 81 to 91% sulfur retention in the bed, with a utilization of 37 to 41%.

The NCB results are summarized in Table 22. Utilization has been calculated from the sulfur retention value and the Ca/S mole ratio in the bed. In Task 1 experiments, the highest utilization observed was 62.5%. This experiment was conducted with fines rate recycle about five times the coal feed rate; the particle size distribution was such that 50% of the dolomite lay in the range below 125 μ diameter particles. This high utilization may be associated with fine particles having an effectively long residence time in the bed.

TABLE 22

CALCIUM UTILIZATION IN FLUIDIZED BED EXPERIMENTS -- NATIONAL COAL BOARD RESULTS WITH DOLOMITE 1337

Task	Test	P/atm	Part Size	S Retention (%)	Ca/S	% Utilization	
						Maximum	Minimum
1	4.1	1.0	1680	85	3.1	27.4	
	4.2	1.0	1680	75	2.6	31.2	
	4.3	1.0	1680	76	2.7	28.2	
	4.4	1.0	1680	75	2.7	27.8	
	4.5	1.0	1680	86	2.2	39	
	4.6	1.0	1680	99	1.6	62.5 (fines recycle)	
	6.1	1.0	3175	67	2.5	40	
	6.2	1.0	3175	88	5.4	16.3	
	6.3	1.0	3175	90	5.3	17.0	
	6.4	1.0	3175	94	5.2	18.1	
	6.5	1.0	3175	89	5.0	17.8	
2	2.1	3.5	1100	92	1.41	65.0	61.
	2.2	3.5	1100	96	1.88	53.2	
	2.3	3.5	1100	97	1.93	51.8	
	2.4	5.0	1100	95	1.88	50.5	
	3.1	5.0	1100	95	1.65	57.5	48.4
	3.2	5.0	1100	95	1.80	52.7	52.6
	4.1	5.0	400	94	1.54	61.0	55.3
	4.2	5.0	400	98	2.02	48.0	45.9
	5.3	5.0	400	85	1.08	78.6	68.4
3	3.7	1.0	<1587	76	1.74	43.6	

In Task II experiments conducted at pressures of 3.5 and 5 atmospheres, the average utilization was 57%. This may be an upper limit since allowance has been made for calcium in the coal. Strict Ca/S ratios of the input stream lead to an average utilization limit 8% lower, or about 50%. Since the S/Ca ratio in the bed, primary cyclone, and secondary cyclone increased in the order .324, .447, .625, the finer particles may contribute significantly to the observed utilization.

The results obtained by Esso (R & E) apparently imply faster rates of reaction at higher utilization than observed in the TG study. Thus, predicted utilization for a residence time of 60 minutes in 2700 ppm SO_2 in a fluidized bed is about 48% with 1100 μ particles. Westinghouse experiments at 2500 ppm SO_2 with 500 μ particles predict 50% utilization in one hour but show a lower rate at this utilization, i.e., .002 min^{-1} in place of approximately .003 min^{-1} shown by the Esso results. (Our data relate to a constant 2500 ppm SO_2 flow over the solid.) However, dolomite 1337 attrited rapidly in these experiments. Using 25/50 mesh 1337 stone, 50% of the material was collected in an adsorption/regeneration cycle in which 52.5 CaO utilization occurred.

Thus, the overall utilization achieved in fluidized beds falls in the range 35 to 50%, and this corresponds to the end of the fast period of reaction observed in our rate studies at atmospheric pressure, and for the bulk of our pressurized runs (10 atmospheres). If this fast reaction mode governs utilization in the fluidized bed, then the possibility exists that the utilization may be doubled. From our pressurized results on both dolomite 1337 and Tymochtee dolomite, it is clear that a utilization of 80 to 90% can be achieved with 500 μ particles.

There are data from fluidized bed investigations which support this idea. Esso (R & E) calcined limestone 1359 under conditions where the CO_2 left the carbonate sites at a dissociation pressure above atmospheric pressure. The resulting calcium utilization in reaction with SO_2 increased three-fold over that observed when calcination occurred under low CO_2 pressures, (6 to 8%) as against (1.5 to 3.5%). Consolidation Coal Co. obtained excellent SO_2 removal (> 90%) in fluidized beds of relatively coarse (16 x 28 mesh) Tymochtee dolomite, at 1800°F. In commenting on the results which are clearly superior to most of the data which has been obtained by other workers, they suggest that rapid but controlled high-temperature calcination is a factor which may contribute to the high sulfur retention at high calcium utilization.

3.10 QUANTITATIVE COMPARISON WITH FLUIDIZED BED RESULTS

The data obtained on the TG apparatus may be used as input to the model developed by Koppel[15] for prediction of sulfur dioxide retention in fluidized beds of limestone or dolomite.

Two cases were considered, one using conditions which obtained during an ANL test with dolomite 1337 for which the experimental result is known, and the other using design data for a projected adiabatic combustor.

The most important parameter in Koppel's model is the "equilibrium utilization," Y_e, achieved with a particular sorbent. The model uses the expression $Y = Y_e\ [1-e^{-kt}]$ to describe the course of sulfation. The rate constant k may be obtained by fitting the TG data to the rate expression. Such curve-fitting involves an arbitrary allocation of Y_e. Figure 77 shows the data for TG 158 curve-fitted in this manner. The deficiency of this curve-fitting procedure is that reaction is not permitted to proceed beyond Y_e, although in practice, there is a finite but small rate of reaction beyond this point.

When the data of TG 108 are applied to Koppel's model, they predict the retention which would be obtained under a given set of conditions. Applying the conditions for the Argonne experiment gives a predicted sulfur retention of 87%. The result obtained by Argonne was 81%, with a CaO utilization of 37%. When a maximum calcium utilization of 37% was used, the model predicted 81% sulfur retention. Clearly the choice of the equilibrium utilization fixes the sulfur retention, given that other conditions remain constant. The quality of the results may be judged on the agreement between the TG data and the course of reaction as shown by the simulation.

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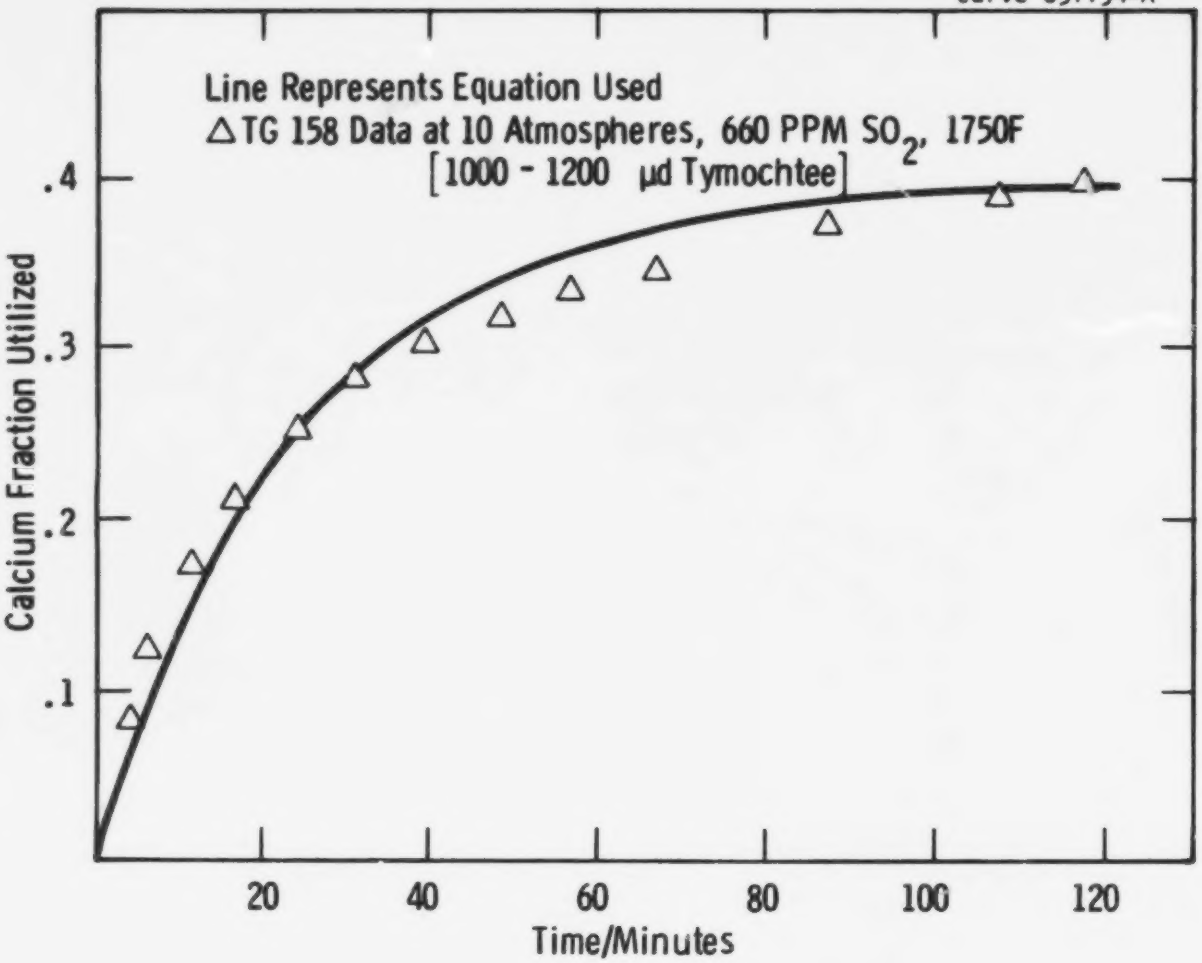

Figure 77: Data Used in Koppel's Model for Sulfation

For the adiabatic combustor, two cases were examined. Uniform SO_2 generation in the fluidized bed was assumed. The normalized bed height required for 90% sulfur retention in the bed was calculated, leading to values for the bed height of 12.7 feet for the eight-module case (64 ft^2 each) and 4.62 feet for the four-module case (352 ft^2 each). The TG data used are shown in Figure 77. The data are pessimistic for two reasons: high utilization was not achieved in this run; and the model assumed no reaction occurs beyond 40% utilization.

TABLE 23

DESIGN CONDITIONS FOR THE ADIABATIC COMBUSTOR
SINGLE BED, DESIGN 1

Item	Condition
Number of Modules	4
Number of Beds per Module	1
Module Diameter, ft	21
Module Height, ft	16
Bed Area, ft^2	347
Fluidizing Velocity, ft/sec	6

30 mole % Ca utilization of the dolomite
90% removal of SO_2
4.3% S in coal
SO_2 level without SO_2 removal = 1070 ppm
Dolomite feed rate -- 179,000 lb/hr
Coal feed rate -- 140,000 lb/hr
Bed height from Koppel's model and TG 158 = 4.6 ft

3.11 COMPARISON OF NCB DATA AND WESTINGHOUSE DATA

- NCB reports a rate constant for 548 μ diameter 1337 dolomite of .112 m^3/kg sec at 4.6% calcium utilization and 1510°F.
- Westinghouse atmospheric data at 1562°F for 450 μ diameter 1337 dolomite at 5% calcium utilization is .45 m^3/kg sec.

TABLE 24

EXPERIMENTS USING TYMOCHTEE DOLOMITE

TG	[SO_2]	P/atm	T°F	Calcination	Mesh	Comment
86	2500	10	1472	N_2		
87	2500	10	1472	N_2		
88	2500	10	Scan	N_2		
89	2500	10	1562	N_2		
90	1250	10	1562	N_2		
97	5000	10	1562	N_2		
98	5000	10	1535	N_2		
147	5000	10	1562	N_2		
153	5000	10	1600	CO_2	5/6	
154	5000	10	1600	CO_2	16/18	
155	5000	10	1700	CO_2	16/18	
156	5000	10	1850	CO_2	16/18	
157	660	10	1750	CO_2	5/6	high O_2 (11%)
158	660	10	1750	CO_2	16/18	11% O_2
159	660	10	1750	CO_2	35/40	11% O_2
160	90	10	1750	CO_2	16/18	10% O_2

Composition: 49.3% $CaCO_3$[a]
36.6% $MgCO_3$

---

[a] Argonne Monthly Progress Report, September 1972.

TABLE 25

EXPERIMENTS WITH DOLOMITE 1337

TG	$[SO_2]$	P/atm	T°F	Calcination	Comment
91	2500	10	1590	N_2	large sample, may hinder kinetics
92	2500	10	1562	"	"
93	1250	10	1562	"	"
94	2500	10	1542	"	"
95	5000	10	1562	"	"
96	2550	10	1620	"	"
102	5000	1	1562	"	
103	5000	1	1562	"	
104	5000	1	1562	"	
105	5000	1	1562	"	
106	5000	1	1562	"	
107	5000	1	1562	"	
108	5000	1	1562	"	
110	5000	10	1562	"	high utilization noted in sulfation
127	5000	10	1562	"	
137	2500	10	1700	"	
138	2500	10	1661	"	
139	5000	10	1724	"	
140	5000	10	1600	"	
141	5000	10	1652	"	
142	5000	10	1652	"	
143	5000	10	1652	"	
145	5000	10	1750	"	
146	820	10	1700	"	
148	5000	10	1353		
149	5000	10	1600		
150	5000	10	1562		high utilization noted

NCB reports that SO_2 stagnation limited the rate to a maximum possible value of .131 m^3/kg sec. It is, therefore, appropriate to compare the rates at higher utilization.

For TG 107 at atmospheric pressure, the rate at 22% utilization is 1.1×10^{-1} m^3/kg sec.

For NCB run 53 at 22.3% utilization, the rate is 9.77×10^{-2} m^3/kg sec.

3.12 REGENERATION

The objectives of sorbent regeneration systems are:

- To reduce the quantities of dolomite or limestone required for sulfur removal, by recirculating regenerated stone
- To reduce the quantity of waste solid rejected from the process
- To recover sulfur, either in useable or storable form.

Assessment of the schemes proposed for reduction of calcium sulfate to calcium oxide or calcium carbonate led to the conclusion that two regeneration processes appeared likely to be economically and technically successful.

In the first of these processes, high-temperature reduction of calcium sulfate to calcium oxide yields sulfur dioxide for recovery:

$$CaSO_4 + \left.\begin{matrix} H_2 \\ CO \end{matrix}\right\} \longrightarrow CaO + SO_2 + \left.\begin{matrix} H_2O \\ CO_2 \end{matrix}\right\} .$$

In the second two-stage process, a low-temperature reduction of calcium sulfate to calcium sulfide is followed by conversion of calcium sulfide to calcium carbonate,

$$CaSO_4 + \left.\begin{matrix} 4H_2 \\ 4CO \end{matrix}\right\} \longrightarrow CaS + \left.\begin{matrix} 4H_2O \\ 4CO_2 \end{matrix}\right\}$$

$$CaS + H_2O + CO_2 \rightarrow CaCO_3 + H_2S,$$

sulfur being recovered by further processing of the hydrogen sulfide stream.

A summary of the apparent advantages and disadvantages of the two processes is shown in Table 26.

The two-stage process was selected for investigation on the TG apparatus. While it is difficult to choose between the two processes, the two-stage process had less experimental data available for an adequate assessment of its potential. Further, its applicability at pressure made it a natural subject for investigation on the TG system. Additionally, the second stage is also under investigation in connection with sulfur removal and sorbent regeneration studies of fluidized bed coal gasification.

Studies on the two-stage process have also been carried out at Argonne. The second step has been investigated by Consolidation Coal Company.

3.13 PREVIOUS REGENERATION STUDIES

Attempts to recover sulfur from calcium/sulfur salts were investigated by Riesenfeld (1920),[22] Rosenquist (1951),[23] and more recently by Squires[24] and his school.

Squires has reported preliminary work on the pressurized conversion of calcium sulfide to calcium carbonate, in which the feasibility of H_2S-rich off-gas streams has been demonstrated. A 24% H_2S gas stream was obtained at 1100°F and 220 psia.[24]

Parallel with the work reported here, Consolidation Coal Company have demonstrated multiple recycling of the reactions

$$CaCO_3 + H_2S \longrightarrow CaS + H_2O + CO_2$$

$$CaS + H_2O + CO_2 \longrightarrow CaCO_3 + H_2S$$

TABLE 26

REGENERATION PROCESSES

	Advantages	Disadvantages
One-Step:	single-stage process	reduction may produce unwanted calcium sulfide in competing reaction, consuming reducing gas
	experimental data available	high temperatures required to avoid problem above, may sinter and hence deactivate the sorbent, and use heat
	CaO formed directly	equilibrium SO_2 concentration decreases with pressure
Two-Step:	thermodynamics favored by low temperature	competing reaction reduces sulfate to yield SO_2 (this disadvantage trades against a sufficiently high temperature to achieve reasonable kinetics)
First Step	no thermodynamic disadvantage of pressure	possibility of melt formation by eutectics in the systems $Ca(OH)_2/CaO$, $CaSO_4/CaS$
	formation of SO_2, COS, H_2S, CS_2, S suppressed at low temperature	two stages required to form $CaCO_3$
	low temperatures avoid solid sintering	
Second Step	pressure thermodynamically favors H_2S production	little experimental data available;requires experimental verification
	continued low temperatures avoid solid sintering	requires high pressure CO_2, H_2O
		produces carbonate rather than oxide
		kinetics and equilibrium H_2S require opposing temperature conditions

in a fluidized bed reactor system at 15 atmospheres.[25] Their results are noted in a later section.

Experimental work at Argonne[26] has yielded results on both the calcium sulfate reduction reaction and on continuous cycling from sulfate to sulfide to carbonate. These results are also noted below.

The experiments on the TG apparatus carried out at 10 atmospheres pressure fell into two parts.

- Experiments in which dolomite was sulfated and the calcium sulfate was reduced to calcium sulfide. In some of these experiments, regeneration of the calcium sulfide to calcium carbonate was attempted.
- Experiments in which calcium sulfide, prepared by direct reaction of hydrogen sulfide with dolomite, was regenerated to calcium carbonate, and experiments using reagent-grade powdered calcium sulfide.

In all the experiments, 35/40 mesh stone was used, with the exception of the cyclic regeneration experiment where 16/18 mesh stone was tested, as noted in Figure 81.

3.14 CONCLUSIONS FROM THE REGENERATION EXPERIMENTS

Experiments on the two-step regeneration led to the following conclusions:

- Calcium carbonate can be regenerated from sulfided limestone and sulfided dolomite.
- Sulfated dolomite can be reduced to dolomitic calcium sulfide at 1600°F.
- Sulfided dolomite produced by reduction of sulfated dolomite is relatively inert to regeneration.

- Inert calcium sulfide gradually builds up in dolomite when it is directly sulfided with hydrogen sulfide and regenerated in cyclic fashion.
- Future studies should concentrate on elucidating the nature of this inactive calcium sulfide.

3.15 TWO-STEP REGENERATION EXPERIMENTS

The first experiments on sulfation regeneration are listed in Tables 27 and 28. Dolomite, in either the calcined or half-calcined state was sulfated on the TG apparatus. The sulfate was reduced to calcium sulfide, and, in some cases, regeneration of the calcium sulfide to calcium carbonate was attempted.

In the first experiment with fully calcined dolomite 1337, sulfated to 33% calcium utilization, reduction in 10% CO at 1382°F gave an apparent 98% yield of sulfide from calcium sulfate. Chemical analysis yielded 87% of the expected calcium sulfide. The reduction time was 140 minutes. In the next experiment, reduction to calcium sulfide was carried out in a $CO/CO_2 = 1/1$ mixture. The rate at 1382°F was slow (16% S reacted per minute), but it increased to (44% S reacted per minute) on heating to 1472°F. To reduce the sulfate present would have taken about four hours in this mixture, at this temperature.

In the next experiment, TG94, after reduction to sulfide under CO_2 (thus maintaining the carbonate phase), the reaction with H_2O and CO_2 was tried at 1200°F (650°C) (Figure 78). A rapid initial weight uptake corresponding to 20% regeneration occurred, followed by an extremely slow reaction ($\sim$ 5% carbonate formed per hour). Increasing the carbon dioxide concentration had no effect. An attempt was made to test the reactivity of the regenerated stone to sulfur dioxide and oxygen. However, exposure of the stone to the gases resulted in parallel reactions, in which the residual sulfide content also oxidized. This was demonstrated by stopping the supply of sulfur dioxide; oxidation proceeded unchecked.

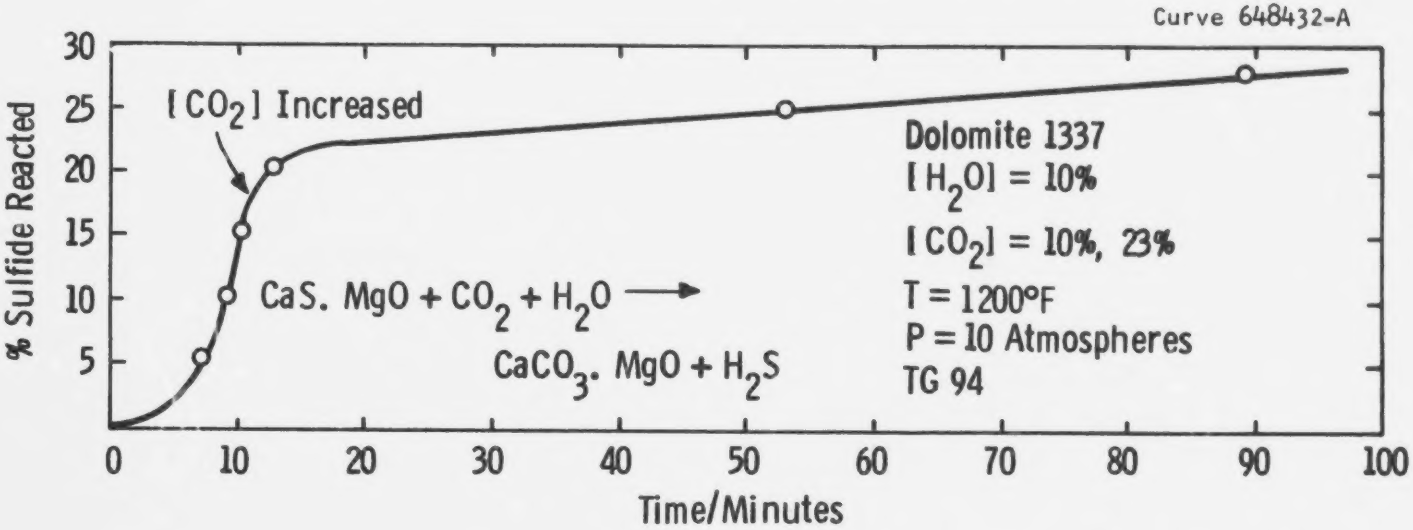

Figure 78: Regeneration of Dolomite

TABLE 27

TWO STEP REGENERATION AT 10 ATMOSPHERES

Experiment Number	Sorbent	% Sulfation	Reduction Conditions	Sulfide Yield	Note
92	calcined 1337	34	10% CO; 1382°F	98%	chemical analysis showed 87% yield of sulfide
93	calcined 1337	27	[CO] = [CO_2] = 10% 750, 1471°F	97%	
94	half-calcined 1337	36	CO/CO_2 = 2/1 CO = 25% 1562°F	100%	
95	half-calcined 1337	30	CO/CO_2 = 2/1 CO = 20% 1508°F	98%	
96	half-calcined 1337	38	CO/CO_2 = 2/1 CO = 20% 1508°F	98%	
97	half-calcined Tymochtee	70	CO/CO_2 = 2/1 CO = 20% 1562°F	97.5%	
98	half-calcined Tymochtee	76.5	CO/CO_2 = 2/1 CO = 20 1562°F	97.5%	

TABLE 28

TWO-STEP REGENERATION AT 10 ATMOSPHERES

Experiment Number	Second-Step Conditions	Temperature	Carbonate Yield	Time (Mins.)	Further Reaction
94	$H_2O = CO_2 = 10\%$	1200°F	28%	180	
95	$CO_2 = 10\%$	1560°F	55%		no reduction observed in CO, so active sulfate not produced by CaS oxidation in CO_2
96	$H_2O = CO_2 = 10\%$	1100°F	13%		some of residual sulfide oxidizable in 4% O_2
97	$H_2O = CO_2 = 10\%$	1560°F	63%		residual S_2^- (by wt) = 1.22×10^{-4} moles Residual S_2^- (titration) = 1.3×10^{-4} moles
98	$H_2O = CO_2 = 20\%$	1100°F	16.5%		

Further experiments compared the reduction of sulfated dolomite at different sulfate levels. The kinetics of reduction for samples sulfated to 26% and 38% calcium utilization were similar, (Figure 79). While weight change measurements invariably indicated total reduction of sulfate to sulfide, some contribution from the parallel reduction to calcium oxide and sulfur dioxide may have been present. Because of the high-volume gas flow maintained over the sample and the dilution effect from the purge nitrogen, it was not feasible to determine the sulfur lost in the gas stream analytically. However, after reduction of the calcium sulfate at 1562°F,and regeneration of the sulfide to carbonate, residual sulfide was determined iodometrically on a portion of the solid product. Sulfide expected from weight change was 1.22×10^{-4} moles. Sulfide by iodometric titration was 1.3×10^{-4} moles. It was concluded that the predominant product in sulfate reduction was calcium sulfide and that stone with a high level of calcium utilization in sulfation could be completely reduced.

Attempts to regenerate carbonate from the calcium sulfide were largely unsuccessful. Stone with a low original calcium utilization in sulfation gave 28% regeneration at 1200°F. Stone with high utilization gave 16% regeneration at 1200°F. The mole-fraction of regenerated calcium carbonate was, however, similar in each case. This suggests a physical mechanism in which a certain amount of carbonate sufficient to seal off transport pores can form, after which reaction is limited by slow diffusion processes. The possibility that the calcium sulfide is reoxidized to calcium sulfate by carbon dioxide, reversing the reduction step, was considered. However, if such sulfate is produced, it is not itself reduced by a carbon monoxide stream. In addition, regeneration at a temperature of 1560°F proceeded to 55 to 63% of the calcium sulfide in TG95 and TG97 demonstrating the possibility of regeneration. However, at such high temperatures, the concentration of sulfur in the product stream would be far too low for recovery purposes.

Additional work concentrated on the second-stage reaction.

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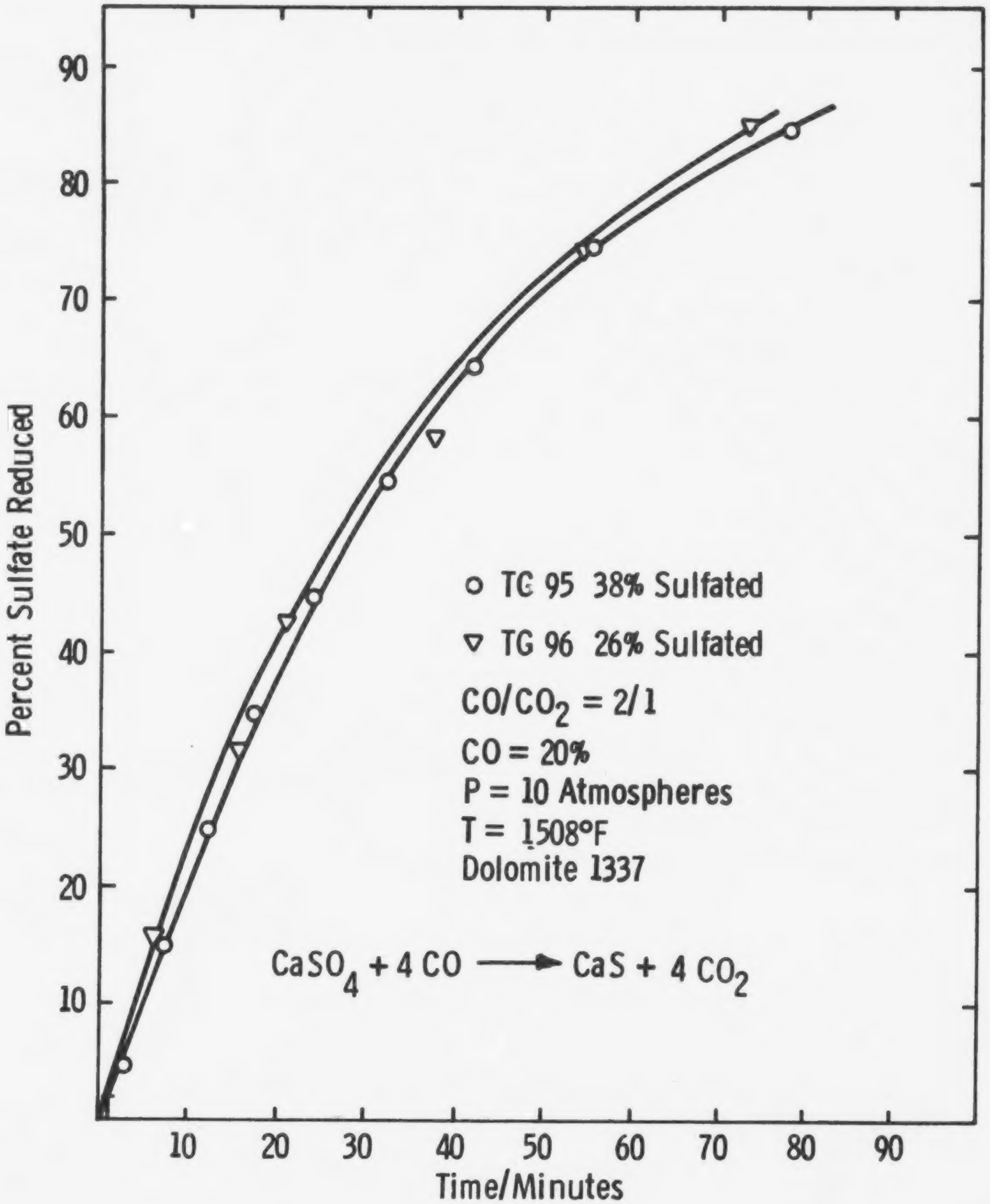

Figure 79: Reduction of Sulfated Dolomite

3.16 EXPLORATORY RUNS

In exploratory runs, a dolomite was sulfided and regenerated at atmospheric pressure, and a limestone was sulfided and regenerated at 10 atmospheres.

The atmospheric pressure experiment involved sulfidation of Tymochtee dolomite in 15% H_2S. In fact, the magnesium content of the stone sulfided, and exposure to nitrogen containing 10% H_2O resulted in rapid hydrolysis of the magnesium sulfide. Some hydrolysis of the calcium sulfide permitted resulfidation, and it appeared that sulfidation was slower on cycles after the first one.

The pressurized experiment, at 10 atmospheres, was carried out with three sulfidation runs and two regenerations in cyclic fashion. The course of reaction observed was as follows:

- 92.5% of CaO reacted to CaS.
- 56.5% of CaS reacted to $CaCO_3$, then to CaO.
- 100% of the CaO converted to CaS.
- 47% of the CaS reacted to $CaCO_3$, then to CaO.
- 100% of the CaO formed converted to CaS.

The rates of sulfidation were the same in each cycle up to 30% utilization.

3.17 TESTS WITH POWDERED CALCIUM SULFIDE

At atmospheric pressure and 1112°F, 8% H_2O and 8% CO_2 achieved 4.1% conversion to carbonate in four hours, and the rate of reaction over the last hour was < 0.2 mole % hr^{-1}. On heating the sample to 1472°F, to measure the carbonate formed, the reaction revived, the rate increasing to 8% hr^{-1}. Apparently, the rate of reaction is very sensitive to temperature in this region. Regeneration was attempted using 50% H_2O, 50% CO_2 at 1112°F. Within two minutes, 2% conversion was attained but the rate fell off to 1% hr^{-1}. The temperature was then increased at a constant heating rate, and the rate of reaction rose to 0.6% min^{-1} at 1562°F. At this temperature, the calcium carbonate decomposed to oxide, and continued conversion of the calcium sulfide to calcium oxide was observed, the rate at 1652°F being 3% min^{-1}. At the start of this experiment, the powdered calcium sulfide was held in steam/N_2 for thirty minutes, and no weight change was observed.

A sample of powdered calcium sulfide was then reacted with 50% H_2O/N_2 at 1562°F; the rate of reaction was 0.8% min^{-1} initially, and 70% conversion was achieved in 133 minutes.

A test at 10 atmospheres on powdered CaS was then carried out at 1292°F. Conversion to carbonate was ∿ 10% in one hour.

It was concluded that the reaction should be studied at the highest temperature consistent with the thermodynamic requirements for the hydrogen sulfide off-gas concentration, or about 1292°F.

3.18 TESTS WITH SULFIDED DOLOMITE

TG 124 was an attempt to regenerate a previously sulfided batch of dolomite. The dolomite contained 1.450×10^{-4} moles of sulfide according to iodometric analysis. The weight gain in the experiment during regeneration indicated that 1.452×10^{-4} moles of carbonate was formed. Decomposition of the carbonate by heating showed that 1.432×10^{-4} moles of carbonate had formed.* It was concluded that 81.3% of the calcium sulfide had been regenerated (Figure 80).

At the end of the experiment, some of the dolomite particles were embedded in plastic resin and ground to expose a cross section through the interior of the stone. The sample was then stained with lead acetate solution. The center of the particle stained heavily, and the outside rim remained white, indicating that the bulk of the residual sulfide was held in the core of the solid.

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*N.B. The remainder of the carbonate formed was the product of calcium oxide recarbonation.

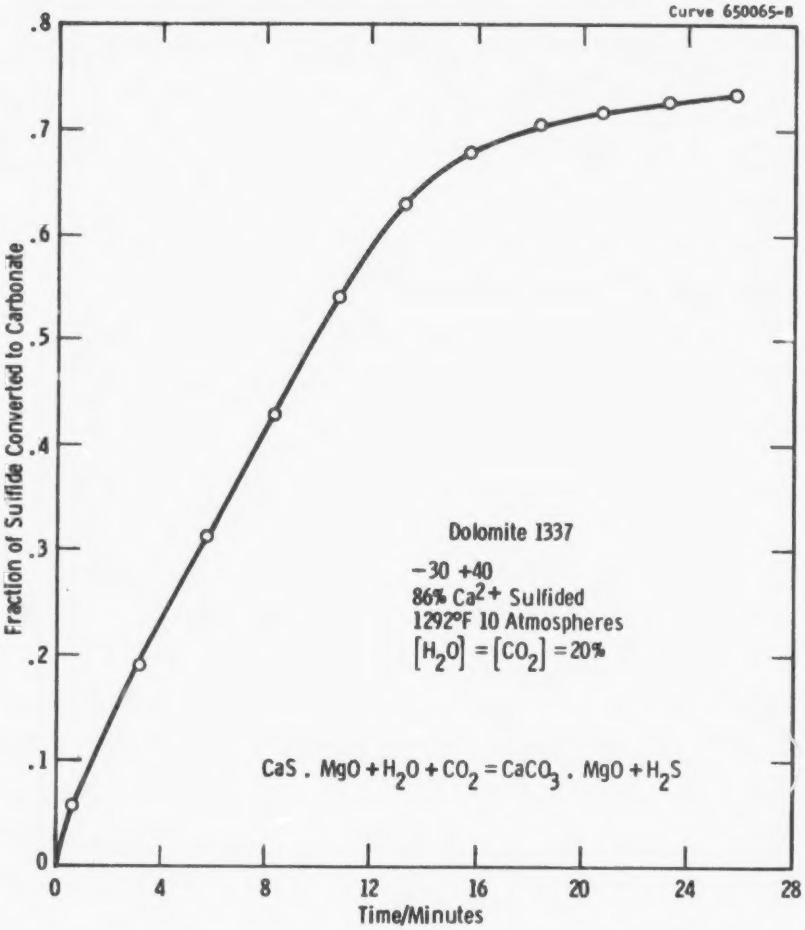

Figure 80: Regeneration of Carbonate from Sulfided Dolomite

The rate of reaction in experiment TG 124 was reproducible, as determined by repetition of the experiment. The rate was in reasonable agreement with that reported by Pell at atmospheric pressure:

0.88 $\sec^{-1}$ Pell

1.23 $\sec^{-1}$ TG 124, TG 125.

For TG 124, assessment of the carbonate formed by reaction was confirmed by decomposing the carbonate thermally. Agreement within 1% was noted.

3.19 CURRENT AND FUTURE EXPERIMENTS

Experiments have been commenced to determine the activity of calcium carbonate formed on regeneration and the number of cycles in which a stone sample is useful for desulfurization.

A sample was sulfided to ∿ 60% sulfidation, resulfided, and regenerated. Figure 81 shows the calcium sulfide content of the stone after the sulfidation step and after the regeneration step. In all, 3.37 moles of sulfur are transferred from desulfurizer to regenerator per mole of calcium in the sample. In each cycle, the regeneration step was continued until the rate became extremely slow; choice of the degree of sulfidation in each cycle was arbitrary. Experiments on this system are continuing.

The utility of these experiments in relation to fluidized bed combustion regeneration systems will depend on the degree of success achieved in identifying inert calcium sulfide. There are three current postulates as to why regeneration does not succeed after reduction of the sulfate:

- The morphology of calcium sulfide produced on reduction of calcium sulfate differs markedly from that produced in the sulfidation of calcium carbonate. This could be a combination of crystallite size and porosity effects.

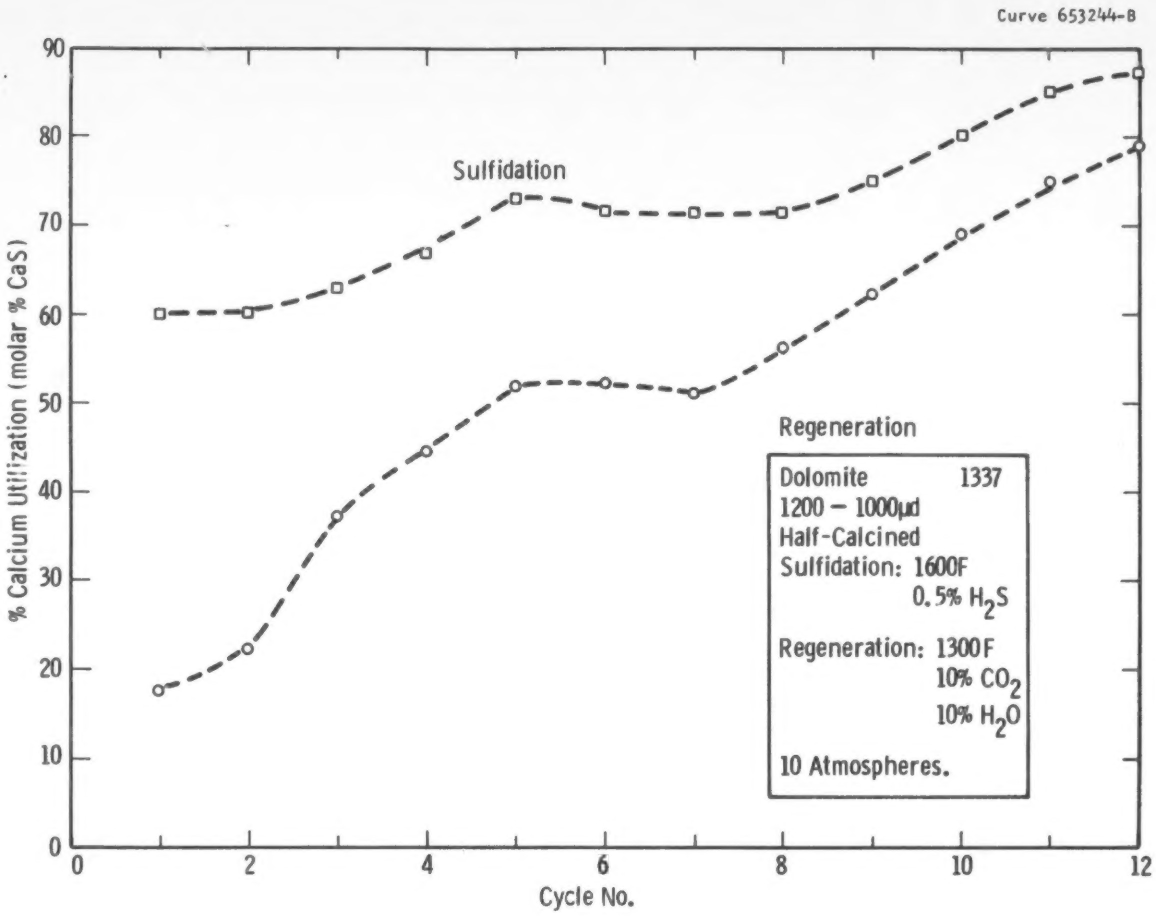

Figure 81: Regeneration of Half-Calcined Dolomite

[Time for Regeneration: 45 Minutes (1st Cycle)
15 Minutes (11th Cycle)]

Examination of samples by electron microscopy and porosity measurements on samples prepared in a fluidized bed should help in determining the influence of this factor.

- Eutectic melting or incipient melting caused by a calcium sulfate/calcium sulfide eutectic during reduction may seal the internal surfaces of the stone. Careful determination of this eutectic and the conditions under which it forms should help to estimate its effect.
- The oxygen activity over the stone during the reduction may establish the structure of a regenerable or a nonregenerable calcium sulfide. Studying the kinetics of regeneration after different reduction treatments would determine this effect.

The current status of other work on this process is summarized as follows:

- Experiments by Consolidation Coal Company on the second stage of the regeneration process have demonstrated regeneration in fluidized bed experiments over a number of cycles (up to 18).[25] The results are similar to those shown in Figure 81, in that there is a continuous build-up of inactive calcium sulfide in the stone.
- Argonne has demonstrated that cyclic two-step regeneration yields only a small amount of regeneration after the first cycle, whether judged by the calcium carbonate produced or the concentration of hydrogen sulfide yielded in the effluent gas stream.[26]

Since TG experiments have yielded the same kind of results that fluidized bed experiments have demonstrated, it is recommended that current experiments to probe the fundamental nature of sorbent inertness to regeneration be pursued.

3.20 CONCLUSIONS

- A thermogravimetric analysis system (TG), modified to operate at high pressures and high temperatures to study the reactions of corrosive gases with solids,
  - Produced quality data on sulfur removal/sorbent regeneration
  - Provided the capability to isolate chemical reactions for study while maintaining close simulation of plant conditions of temperatures, pressure and gas composition
  - Provided a reliable and efficient method for obtaining design data.
- High utilization of calcium in limestones/dolomites for sulfur removal can be achieved by adjusting the carbon dioxide partial pressure during calcination. The technique can be applied to a pressurized fluid bed combustion system to approach the stoichiometric requirement of sorbent stone in a once-through system, thus reducing the tonnage of by-product solids and raw stone requirements.
- The investigation of the regenerative process using the pressurized TG system yields results in close agreement with those obtained in large-scale test units. None of the available data, to date, show adequate regeneration under conditions suitable for sulfur recovery.
- Use of the thermogravimetric analysis systems to investigate the fundamental chemical problems in the regeneration systems offers the possibility of improving these processes for commercial application.

- Data show that TG experimental results can be accurate predictors for kinetic limits of sulfur removal in the pressurized fluidized bed boiler system

4. REGENERATION SYSTEMS ASSESSMENT

4.1 INTRODUCTION AND SCOPE OF INVESTIGATION

The operation of the pressurized fluid bed combustion process with coals presently available to most utilities in the U.S. results in the production of large quantities of partially utilized dolomite or limestone materials in the form of $CaSO_4$. This sorbent material may be either regenerated to a form suitable for repeated SO_2 removal in the fluid bed boiler or disposed of in its partially utilized form in a once-through system.

Regeneration processes are presently under investigation by Esso Research and Engineering Co., Argonne National Laboratory, and Westinghouse Electric Corporation to determine the specific nature and behavior of various processes on a small scale. The question of whether or not any of the regeneration processes currently being investigated are economically and environmentally attractive on a commercial scale, based on the results of small-scale experimental work, should be dealt with before any process is operated on a large scale. Dealing with this question should indicate the relative merits of the once-through sorbent operation and should also indicate the need to pursue new regeneration concepts.

In the present work, the one-step and two-step regeneration processes currently being considered are studied. The design, performance, and cost of a commercial utility power plant utilizing pressurized fluid bed combustion with sorbent regeneration is projected. The potential problem areas and the most sensitive cost factors are established for each of the regeneration processes. The fluid bed power plants utilizing the regeneration processes are compared to the once-through operation of the fluid bed combustion power plant, a fluid bed combustion plant utilizing the Wellman-Lord stack gas cleaning process, and a conventional coal-fired power plant utilizing limestone wet scrubbing or the Wellman-Lord stack cleaning process for pollution control.

The study selects base designs for each of the regeneration processes from the available experimental information. Optimization or detailed equipment design is not undertaken. Standard methods for preliminary cost estimations are applied.

4.2 REGENERATION PROCESS CONCEPTS

The economics and performance of three regeneration processes which function to regenerate the utilized SO_2 sorbent produced in the boilers of the pressurized fluid bed combustion power plant, have been projected. These processes are a one-step process operated at 10 atmospheres pressure, the same one-step process operated at 1 atmosphere pressure, and a two-step regeneration process. A sensitivity analysis has indicated the potential of the one-step process at both high and low pressures (Appendix L). In addition to the consideration of these continuous regeneration process operations which follow the variable power plant load, the concept of a constant load regeneration process has been examined. In this constant load case, storage capacity is situated between the boilers and the regeneration process to allow the regeneration process to operate continuously at a fixed sulfur load, or with on-and-off control. Analysis of once-through operation in which the utilized sorbent is disposed of without further processing has also been carried out to weigh the economic advantages and disadvantages of sorbent regeneration.

4.2.1 One-Step High-Pressure and Low-Pressure Regeneration Concepts

The one-step dolomite/limestone regeneration process consists of a single fluidized bed reaction vessel in which utilized dolomite/limestone ($CaSO_4$) from the pressurized fluid bed boiler (coal-fired) is reacted with a H_2/CO reducing gas to produce CaO and SO_2. The reaction

$$CaSO_4 + \left\{\begin{matrix}H_2\\CO\end{matrix}\right\} \rightleftarrows CaO + \left\{\begin{matrix}H_2O\\CO_2\end{matrix}\right\} + SO_2 \tag{1}$$

requires temperatures on the order of 2000°F to produce high levels (1 to 15%) of SO_2 at reasonable rates. The SO_2 equilibrium concentration is favored by reduced pressure, being inversely proportional to the total pressure. The competing reaction

$$CaSO_4 + 4\begin{Bmatrix} H_2 \\ CO \end{Bmatrix} \rightleftarrows CaS + 4\begin{Bmatrix} H_2O \\ CO_2 \end{Bmatrix} \qquad (2)$$

also occurs to produce the unwanted product CaS.

The regenerated dolomite/limestone is returned to the fluid bed boiler with the required amount of fresh dolomite/limestone to supplement the regenerated sorbent's reduced activity. The SO_2 stream from the regenerator is sent to a sulfur recovery plant in which sulfuric acid or elemental sulfur is produced.

The one-step regeneration process is broken down into five interrelated elements: a regenerator element, a reducing-gas producer element, a sulfur recovery element, a Claus plant tail gas handling element, and a sorbent circulation element. These elements and their relationship to one another are shown in Figure 82. The high- and low-pressure processes are conceptually identical. The simple block flow diagram indicates the basic streams in the process. Fuel, air and/or steam is converted to a H_2/CO reducing gas in the reducing-gas producer element. This reducing gas is combined with tail gas recycled from the sulfur recovery element to make up the reducing gas provided to the regenerator. Utilized sorbent from the fluid bed boiler is transported to the regenerator element to produce a regenerated sorbent and an SO_2 stream. The regenerator product gas is processed in the sulfur recovery element, with a portion of the sulfur recovery tail gas being recycled to the regenerator and the remainder processed in the tail gas handling element. The performance and design of these five regeneration process elements are interrelated in a complex manner.

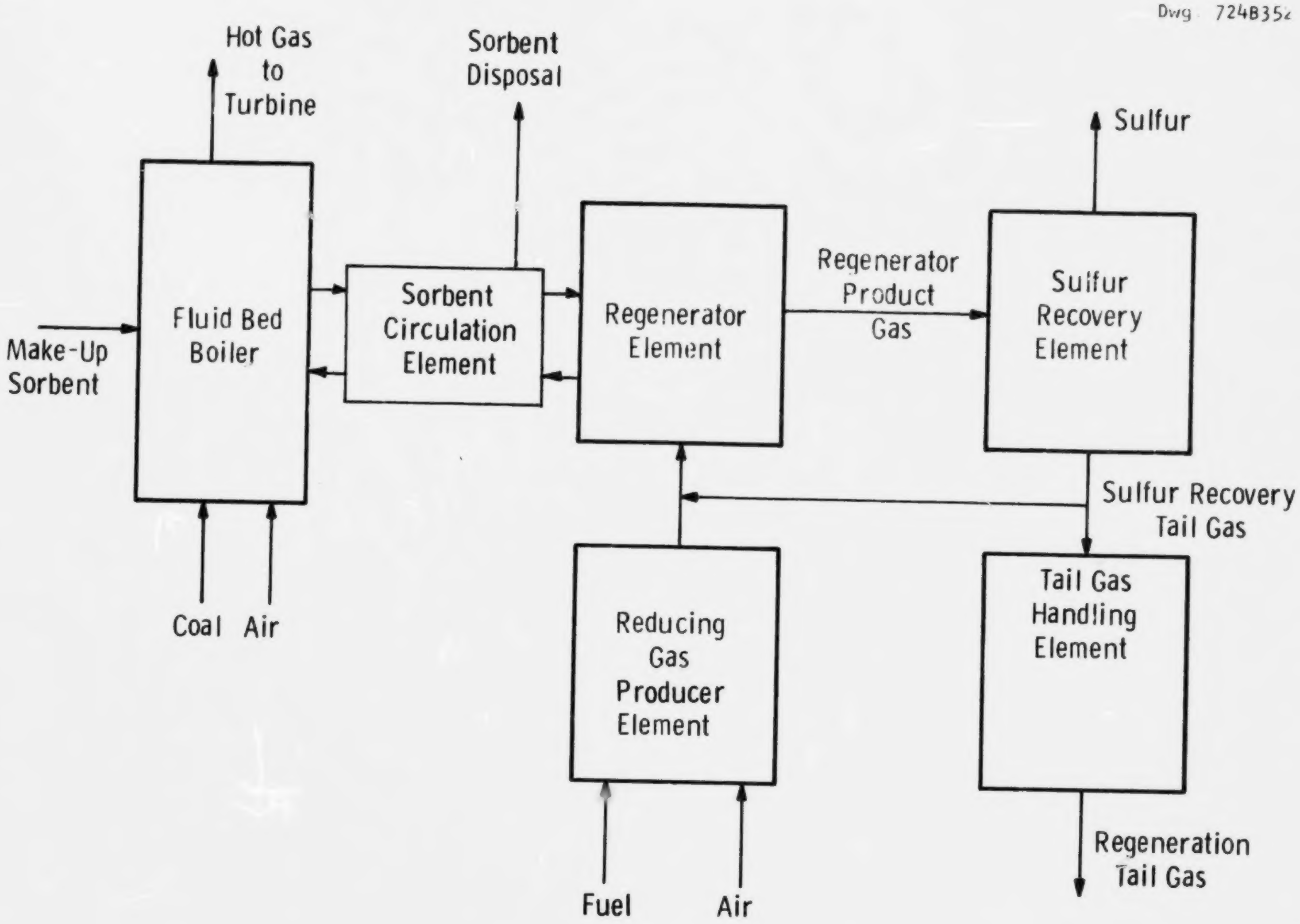
Dwg. 724B352
Hot Gas to Turbine
Sorbent Disposal
Sulfur
Make-Up Sorbent
Fluid Bed Boiler
Sorbent Circulation Element
Regenerator Element
Regenerator Product Gas
Sulfur Recovery Element
Sulfur Recovery Tail Gas
Tail Gas Handling Element
Reducing Gas Producer Element
Coal
Air
Fuel
Air
Regeneration Tail Gas

Figure 82: One-Step Regeneration Process Elements

4.2.2 Two-Step Regeneration Concept

The two-step sorbent regeneration requires two fluidized bed reactors in series to carry out the conversion of calcium sulfate to calcium carbonate. In the first step calcium sulfate is reduced to calcium sulfide by the reaction

$$\mathrm{CaSO_4} + 4\left\{\begin{matrix}\mathrm{CO}\\ \mathrm{H_2}\end{matrix}\right\} \rightleftarrows \mathrm{CaS} + 4\left\{\begin{matrix}\mathrm{CO_2}\\ \mathrm{H_2O}\end{matrix}\right\} \tag{3}$$

at about 1500°F and 10 atmospheres. The second step converts calcium sulfide to calcium carbonate by the reaction

$$\mathrm{CaS} + \mathrm{H_2O} + \mathrm{CO_2} \rightleftarrows \mathrm{CaCO_3} + \mathrm{H_2S} \tag{4}$$

at about 1250°F and 10 atmospheres. The regenerated sorbent is recycled to the boilers while the H_2S stream is sent to a sulfur recovery plant. The two-step regeneration process provides a low temperature route to regenerated sorbent.

Figure 83 illustrates the elements of the two-step process. The process consists of seven elements: the calcium sulfate reducer element, the H_2S generator element, the reducing-gas producer element, the CO_2 recovery element, the sulfur recovery element, the tail gas handling element, and the sorbent circulation element. Reducing gas for the first-step sulfate reduction is provided by the reducing-gas producer element, while CO_2 for the second-step calcium carbonate production is provided by recovering CO_2 from stack gas and recycled sulfur recovery tail gas. Tail gases from the calcium sulfate reducer element and the CO_2 recovery element are incinerated or recycled to the fluid bed boilers in the tail gas handling element. The sorbent circulation element carries out the function of circulating the sorbent between the fluid bed boilers and the regeneration process. Elemental sulfur is recovered in the sulfur recovery element.

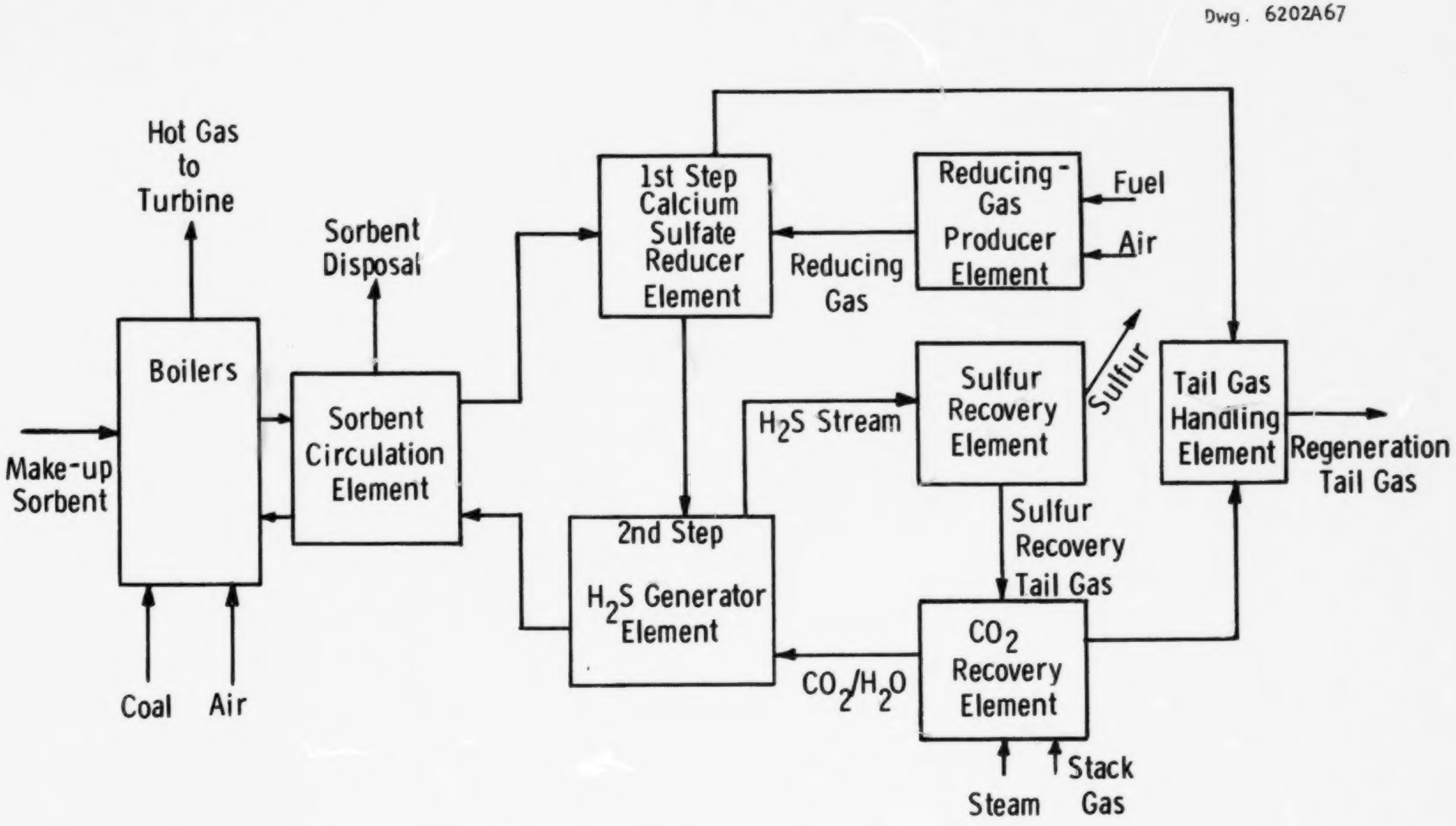
Dwg. 6202A67
Hot Gas to Turbine
Sorbent Disposal
Make-up Sorbent
Boilers
Coal
Air
Sorbent Circulation Element
1st Step Calcium Sulfate Reducer Element
Reducing Gas
Reducing - Gas Producer Element
Fuel
Air
2nd Step H_2S Generator Element
H_2S Stream
Sulfur Recovery Element
Sulfur
Sulfur Recovery Tail Gas
CO_2/H_2O
CO_2 Recovery Element
Steam
Stack Gas
Tail Gas Handling Element
Regeneration Tail Gas

Figure 83: Two-Step Regeneration Process Elements

4.2.3 Constant Load Concept

In order to eliminate the problems involved with designing the regeneration process to follow the power plant load, storage of the utilized sorbent and the regenerated sorbent between the fluid bed boilers and the regeneration process may be used to allow the regenerating process to be operated at a constant load. This concept is shown in Figure 84. Solid sorbent heat exchangers would be required to insure operability and to avoid large thermal efficiency losses.

The concept could be applied to either the one-step or two-step regeneration. If the power plant load fluctuates over a short time period (say one week) then sufficient storage capacity can be supplied so that the regeneration process can be designed at a capacity equal to the average plant load rather than the maximum plant load required in the variable load concept. Such capability can lead to a large cost reduction in the cost of the regeneration process. Fluctuations in the power plant load may occur to the extent that the regeneration process must be designed for maximum plant capacity.

4.2.4 Once-through Operation

Once-through operation with pressurized fluid bed combustion is a simplified operation because the calcium sulfate produced in the boiler is in a stable form suitable for disposal without further processing. The cost of the process is minimized by operating with maximum sorbent utilization. Problems of control are minimized by going to the once-through operation.

4.2.5 Wellman-Lord Stack Gas Cleaning Process

The pressurized fluid bed combustion power plant can also be operated without including a sulfur sorbent in the fluid bed boiler and instead using a final cleaning of the plant stack gas with the Wellman-Lord process. The Wellman-Lord process is a regenerative process producing elemental sulfur and has been successfully operated on a large scale.

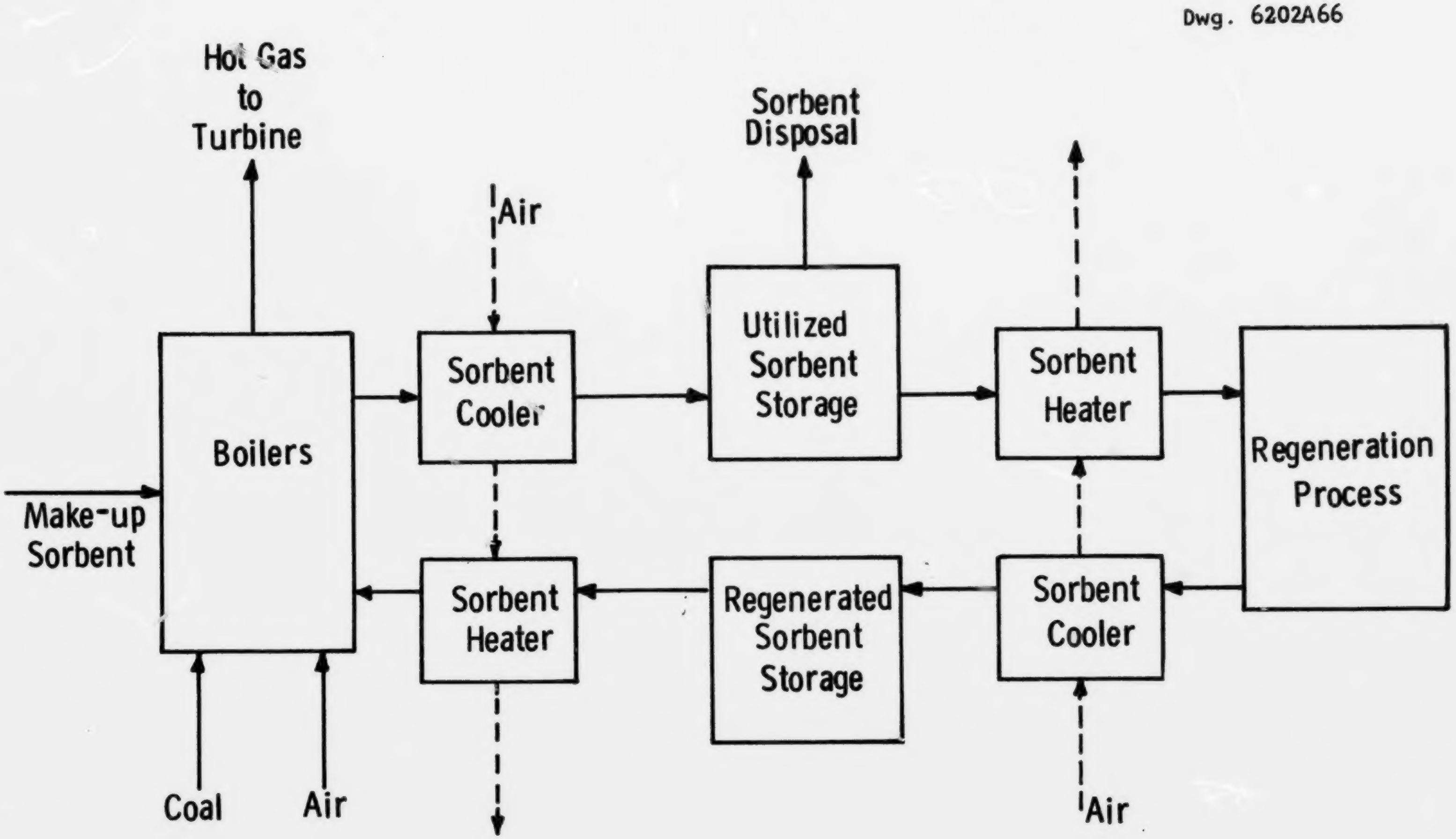
Dwg. 6202A66
Hot Gas to Turbine
Make-up Sorbent
Boilers
Coal
Air
Air
Sorbent Cooler
Sorbent Heater
Sorbent Disposal
Utilized Sorbent Storage
Regenerated Sorbent Storage
Sorbent Heater
Sorbent Cooler
Air
Regeneration Process

Figure 84: Constant LoadConcept

Sorbent regeneration and disposal would be eliminated by this concept at the cost of the problems associated with stack gas cleaning. Economic and environmental projections using this concept are summarized in Section 4.7.

4.2.6 Alternative Concepts

The desulfurization concepts considered for pressurized fluid bed combustion in this study are not meant to represent an exhaustive collection of alternatives. Many other alternatives exist. Sulfur sorbents other than limestone or dolomite might improve the process performance, economics, and environmental impact. The possibility of alternative regeneration processes for dolomite or limestone also exists. These would include both high-temperature and low-temperature concepts. Scrubbing systems other than the Wellman-Lord process could be considered to operate with the pressurized fluid bed power plant. The scrubbing process could also be operated at pressure to desulfurize the cooled gas before it enters the gas turbine. Also, other plant cycles might be considered. These alternatives, and others, should be dealt with in some detail to assess their potential value.

4.3 PROCESS OPTIONS AND VARIABLES

The design options and process variables which are of importance to the proposed concepts in Figures 82 and 83 are listed below.

PROCESS OPTIONS

- Reducing-gas producer fuel and process
- Source and process for CO_2 production
- Recovery of elemental sulfur or sulfuric acid
- Recycle of process tail gas to boilers or incineration
- Limestone or dolomite as sorbent
- Disposal of spent sorbent before or after regeneration
- Number of regenerator modules

Reducing-gas for the reducing stages of the one-step or two-step regeneration processes can be produced by gasification; partial oxidation; or reforming of coal char, residual oil, distillate fuel oils, or methane. These fuel options are examined in Appendix L. The source of CO_2 for the second stage of the two-step regeneration process is also a process consideration having many options. Recovery of elemental sulfur or sulfuric acid from the SO_2 or H_2S streams produced in the processes are additional options. The tail gas from the regeneration process must be either incinerated and released to the atmosphere or recycled to the fluid bed boilers. Incineration might be followed by energy recovery (turbine expander or waste heat boiler) and final clean-up of SO_2. Both limestone and dolomite are candidates for sulfur sorbents, each one having possible advantages and disadvantages. The spent sorbent could be disposed of prior to or following the regeneration step. The choice of disposal location (before or after regeneration) can greatly affect the cost and performance of the regeneration process. The number of regenerator modules in the plant depends on the plant turndown requirements and the capabilities of the various elements to operate at reduced load.

The items in the following list represent a simplification of the complete set of variables which describe the regeneration process. Those variables having the greatest effect on the process design and performance are included.

PROCESS VARIABLES

- Regenerator vessel temperature
- Regenerator vessel fluidization velocity
- Mole fraction SO_2 or H_2S from regenerator
- Reducing-gas composition to regenerator
- Weight fraction sulfur in the boiler coal, W_s
- Sorbent make-up rate, m, moles Ca/mole S
- Boiler sulfur removal efficiency, ε
- Sorbent utilization in the boilers, X_S^B, and after regeneration, X_S^R
- Sorbent purchase plus disposal cost, \$/ton

- Boiler conditions: temperature, pressure, excess air, calcination of the sorbent or carbonization
- Process sulfur load: pounds of sulfur processed in the regeneration process per pound of coal fed to the boilers
- Process sulfur load = $W_s\ (\varepsilon - mx_s^B)$ for sorbent disposal before regeneration
- Process sulfur load = $W_S\ (\varepsilon - mx_S^R)$ for sorbent disposal after regeneration

4.4 BASE CASE DESIGNS

Parametric studies of the economics of one-step dolomite regeneration have considered a wide range of operating and design variables. These studies are included in Appendix L. In order to gain clarity and obtain the most probable regeneration process configuration and cost, base designs of the alternative regenerating processes described for use with pressurized fluid bed combustion have been prepared on the basis of the best existing experimental information.

The design specifications are outlined below:

DESIGN SPECIFICATIONS

- Fixed coal rate to boilers - 430,000 lb/hr (∿ 635 MW)
- Sorbent type - dolomite
- Boiler sulfur removal - 95%
- Variable process sulfur load - 0.01-0.06
- Recycle of process tail gas to boilers or incineration
- Recovery of elemental sulfur
- Reducing-gas production from coal only (methane for sulfur recovery)
- Four regenerator modules (one for each boiler module)

The designs have been carried out for a fixed coal rate to the boilers rather than a fixed plant MW capacity. Dolomite is considered rather than limestone. High levels of sulfur removal are specified (though they may not be required to meet SO_2 emission regulations with all coals). Variable process sulfur loads are considered (See listing in

TABLE 29

DESIGN ASSUMPTIONS

Process	Case	
	10 Atm	1 Atm
One-Step Dolomite Regeneration		
Regenerator pressure, atm	10	1
Regenerator temperature, °F	2000	1900
SO_2 mole fraction, %	1 and 2	10
$CO + H_2$ mole fraction in reducing gas, %	1.5 and 3	18
Sulfur recovery efficiency, %	95	95
Dolomite utilization in boiler, %	30	30
Dolomite utilization after regeneration, %	10	10
Dolomite make-up rate, moles Ca/mole S	1	1
Fluidization velocity, ft/sec	5	5
Boiler conditions	calcination	calcination
Dolomite temperature to regenerator °F	1200	1200
Calcium sulfide in regenerated sorbent, %	0	0
Two-Step Regeneration		
$CaSO_4$ reducer pressure, atm	9	
$CaSO_4$ reducer temperature °F	1500	
H_2S generator pressure, atm	11	
H_2S generator temperature, °F	1250	
H_2S mole fraction, %	3	
$CO + H_2$ mole fraction in reducing gas, %	30	
Sulfur recovery efficiency, %	95	
Dolomite utilization in boiler, %	30	
$CaSO_4$ reduced in reducer vessel, %	100	
Dolomite utilization after H_2S generator, %	10	
Dolomite make-up rate, mole Ca/mole S	1	
Fluidization velocity, ft/sec	5	
Boiler conditions	carbonation	
Dolomite temperature to $CaSO_4$ reducer, °F	1200	

Section 4.3). The options of recycling the process tail gas to the boilers or incinerating the tail gas are considered. Recovery of elemental sulfur is specified, and coal is assumed to be the only source of reducing gas (except for the methane needed for sulfur recovery or incineration). CO_2 is recovered from stack gas and sulfur recovery tail gas in the two-step process.

Design assumptions are listed in Table 29. For the one-step regeneration process the SO_2 mole fractions and the reducing gas requirements are projected from Esso's data. Two SO_2 mole fractions are considered in the high-pressure case. A temperature of 1900°F is used in the low-pressure regenerator based on the difficulty of temperature control if the temperature were set at 2000°F. Boiler conditions have been assumed to be calcination conditions with the one-step regeneration process and carbonation conditions with the two-step regeneration process. Either condition may exist in the boiler depending on the boiler pressure, temperature, and excess air. The boiler conditions chosen for the two regeneration processes are the optimum boiler conditions for the respective process. If the one-step regeneration processes must handle carbonated dolomite or if the two-step process must handle calcined dolomite, then the complexity and cost of the processes will increase drastically. The details of the base designs are presented in Appendices D, E, and F.

The basis on which energy costs were computed is presented below:

COST BASIS

- Mid-1973 costs
- Capital charges at 15%/yr
- O&M at 5%/yr
- 70% capacity factor
- No credit for recovered sulfur
- Coal at 45¢/MM Btu
- Methane at 80¢/MM Btu
- Operating water at 8¢/M gal.

4.5 ECONOMIC RESULTS FOR BASE CASE DESIGNS

Capital and energy costs for the four variable load base designs are considered and the effect of credit for recovered sulfur is indicated. The costs of the constant load concept applied to each of the regeneration processes are projected. The assumption of the state of the sorbent in the boiler (calcined or carbonated) is shown to be a critical factor in the cost and performance of a given regeneration process. Finally, the plant energy costs with the regeneration processes are compared to the plant energy cost with once-through operation, as a function of the sorbent cost.

4.5.1 Process Capital Investment

Results of capital cost and energy cost estimates are shown in Tables 30 and 31. Table 30 gives the capital investment in \$/kw as a function of the process sulfur load for the three regeneration processes. The two-step regeneration process requires the greatest investment of the processes considered. The sensitivity of the investment to the SO_2 mole fraction in the regenerator product is indicated by the results for the one-step process at 10 atmospheres. The investments are comparable to investments for a limestone wet-scrubber system (See Appendix N).

4.5.2 Plant Energy Costs

Energy costs for the total power plant are given in Table 31 as a function of the process sulfur load. The table indicates that the energy cost of the one-step processes at 1 atmosphere and 10 atmospheres (2% SO_2) are about the same and the energy costs of the one-step process at 10 atmospheres (1% SO_2) and the two-step process are about the same. Again, the sensitivity of the energy cost to the SO_2 mole fraction is indicated by the 1 and 2% SO_2 costs for the 10 atmosphere one-step process. The \$2/ton dolomite cost assumed in the table may not be a realistic cost

TABLE 30

CAPITAL INVESTMENT FOR REGENERATION PROCESSES - BASE DESIGNS ($/kw)[a]

Process Sulfur Load	One-Step Process-1 Atm	One-Step Process-10 Atm		Two-Step Process[c]
	10% SO_2[b]	1% SO_2[c]	2% SO_2[c]	
0.01	13.4	18.1	11.7	21.0
0.0	27.9	36.8	24.6	46.0
0.06	47.6	60.6	40.9	76.7

---

[a] Does not include interest during construction and fee - mid-1973 costs
- 635 MW (e) plant capacity.

[b] Incineration of process tail gas.

[c] Recycle of tail gas to fluid bed boilers.

but is not of importance when comparing regeneration processes having identical make-up rates. The figures in Table 31 do not include a credit for recovered sulfur. Table 32 indicates what this credit would amount to as a function of the process sulfur load and the sulfur cost.

Cost breakdowns for the regeneration processes are presented and discussed in Appendices D, E, and F.

4.5.3 Constant Load Concept Economics

The capital investments for the constant load concept applied to each of the regeneration processes is shown in Figure 90. It has been assumed that the fluctuations in power plant load occur over a short time period and are reproduced throughout the year so that the design capacity of the regeneration process can be proportional to the plant capacity factor. A capacity factor of 70% has been assumed. Details of the design are shown in Appendix J.

The energy cost corresponding to each of the constant load regeneration processes is shown in Table 34. The plant energy cost is substantially increased over the variable load costs in Table 31 for the one-step process at 10 atmospheres and the two-step process. The energy cost of the low-pressure one-step process is only slightly increased over the variable load concept energy cost. The advantages to be gained from constant load operation may overweigh this small increase in cost.

4.5.4 Effect of Boiler Conditions

The base regeneration process designs have been evaluated with the assumption applied that the boiler conditions have provided the optimum sorbent conditions entering the regeneration process. This means specifically calcium oxide entering the one-step regeneration process (calcination in the boilers) and calcium carbonate entering the two-step process (carbonation in the boilers). Since this optimum behavior may not occur during actual plant operation, it is important to note the effect of the opposite boiler conditions on the process cost and performance. Conservative process assumptions are applied to obtain the results discussed.

TABLE 31

ENERGY COST FOR POWER PLANT - BASE DESIGNS (mills/kWh)[a]

Process Sulfur Load	One-Step-1 Atm $10\%\ SO_2$	One-Step-10 Atm		Two-Step
		$1\%\ SO_2$	$2\%\ SO_2$	
0.01	11.9	12.3	11.9	12.4
0.03	12.7	13.5	12.8	13.5
0.06	13.8	15.1	14.0	15.2

---

[a] Based on:
- Dolomite purchase and disposal cost of $2/ton
- Capital charges of 15%/yr
- O&M of 5%/yr
- 70% capacity factor
- No credit for recovered sulfur
- Coal at 45¢/MM Btu
- Methane at 80¢/MM Btu

TABLE 32

CREDIT FROM SULFUR SALES (mills/kWh)[a]

Process Sulfur Load	Credit From Sulfur Sales, mills/kWh		
	Sulfur at $10/ton	$20/ton	$30/ton
0.01	0.03	0.06	0.10
0.03	0.10	0.19	0.29
0.06	0.19	0.39	0.58

---

[a] Based on 95% sulfur recovery efficiency and 635 MM(e) plant capacity.

TABLE 33

CAPITAL INVESTMENT FOR CONSTANT LOAD CONCEPT ($/kw)[a]

Process Sulfur Load	Investment. $/kw I			
	One-Step-1 Atm 10% SO_2 [b]	One-Step-10 Atm		Two-Step [c]
		1% SO_2 [c]	2% SO_2 [c]	
0.01	15.1	22.1	17.6	24.2
0.03	29.9	45.8	37.5	52.9
0.06	48.9	73.4	59.0	86.8

[a] Does not include interest during construction and fee -- 635 MW(e) plant capacity, mid-1973 costs, 70% capacity factor.

[b] Incineration of process tail gas.

[c] Recycle of process tail gas to fluid bed boilers.

TABLE 34

ENERGY COST FOR CONSTANT LOAD CONCEPT (mills/kWh)

Process Sulfur Load	One-Step-1 Atm 10% SO_2	One-Step-10 Atm		Two-Step
		1% SO_2	2% SO_2	
0.01	12.0	12.4	12.2	12.4
0.03	12.8	13.8	13.3	13.8
0.06	13.9	15.6	14.6	15.5

Some reduction in cost may be possible by improving the process schemes.

If calcination occurs in the fluid bed boilers and the two-step regeneration process is used for sorbent regeneration, then extensive carbonation will occur in the H_2S generator vessel. The exothermic nature of the carbonation reaction will disrupt the H_2S generator temperature control and the requirement for maximum H_2S concentrations in the product gas. Thus, the carbonation of the sorbent must be carried out in a separate vessel prior to the sorbent regeneration. Carbonation may be carried out by recycling plant stack gas. Table 35 shows the effects of calcination boiler conditions on the energy cost of the two-step process. The results are sensitive to the rate of sorbent circulation and the extent of calcination. Details are discussed in Appendix H. Table 35 indicates an extremely large increase in the plant energy cost compared to the results for carbonation boiler conditions in Table 31.

If carbonation of the sorbent occurs in the fluid bed boilers and the one-step regeneration process is used for sorbent regeneration (high-pressure or low-pressure operation), then extensive calcination will occur in the one-step regenerator vessel. The endothermic nature of the calcination reaction will make temperature control and maintenance of high SO_2 concentration impossible. Thus, calcination must be carried out in a separate vessel prior to regeneration. Heat for the calcination must be supplied by the combustion of coal in a separate coal combustor. The fuel required for this process step will normally be some large fraction of the fuel rate to the fluid bed boilers (depending on the sorbent circulation rate and the degree of carbonation in the boilers) so desulfurization of the hot product gas from the calcination vessel and recovery of the hot gas energy is required. Due to the large scale of this calcination process, the regeneration process would have to become an integral part of the power generation system. This would require major modifications to the present fluidized bed combustion plant design concept and would greatly complicate the plant operation. Details are considered in Appendix H.

Since the fluid bed boiler conditions (temperature, pressure, excess air) change during turndown of the power plant, the sorbent may

TABLE 35

EFFECT OF CALCINATION BOILER CONDITIONS ON TWO-STEP REGENERATION PROCESS[a]

Process Sulfur Load	Increase in Capital,[b] $/kw	Increase in Energy Cost,[b] mills/kWh	Energy Lost,[c] % of Fuel Input	Total Energy Cost, mills/kWh
0.01	9.5	0.36	1.1	12.6
0.03	20.4	0.77	3.2	14.7
0.06	32.9	1.24	6.5	17.7

[a] Basis: $X_S^B = 0.30$, $X_S^R = 0.10$, 100% calcination of regenerated dolomite.
[b] Based on 635 MW of plant output.
[c] Total energy cost in mills/kWh is based on this efficiency loss.

switch between calcined and carbonated states during turndown. This is an additional operating complication to be considered for regeneration processes. The boiler condition (calcination or carbonation) is a critical factor to consider when comparing regeneration processes. The state of the sorbent has a small effect upon the energy cost of once-through operation (See Appendix I).

4.5.5 Process Cost Comparison

The energy costs of the four base regeneration process designs and once-through operation are compared in Figures 85 and 86 as a function of the dolomite purchase plus disposal cost. The figures consider 95% sulfur removal of a 4 wt % sulfur coal. A dolomite utilization of 30% in the boiler and a regenerative make-up rate of 1 mole Ca/mole S are considered. Once-through make-up rates of 2 and 3 moles Ca/mole S are shown. Figure 85 indicates the process costs for the case of sorbent disposal from the process before the sorbent is regenerated, while Figure 86 represents the case of sorbent disposal after the sorbent has been regenerated. The energy cost basis listed in Section 4.4 has been applied to the base conditions also listed there and in Table 30 to obtain the results.

The case requiring sorbent disposal after regeneration yields higher energy costs than the case requiring disposal before regeneration. Disposal before regeneration gives smaller process sulfur loads than disposal following regeneration. The extent of this effect on the plant energy cost depends on the sorbent utilization before and after regeneration (See listing of variables in Section 4.3). The chemical nature of the disposed sorbent wanted is a further consideration. The figures indicate a large cost advantage for once-through operation over the regeneration processes considered even for high dolomite costs. Once-through make-up rates less than 2 moles Ca/mole S may be feasible, leading to greater cost advantages. The effect of boiler conditions must also be considered. The plant operation is also simplified with

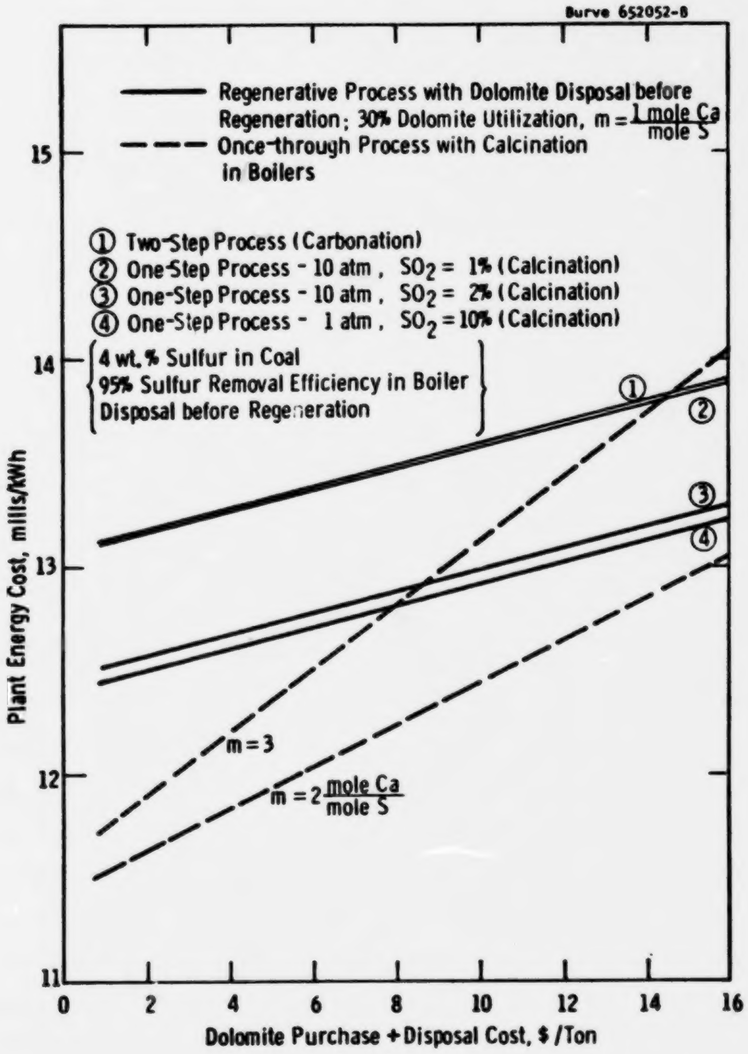

Figure 85: Plant Energy Cost (Disposal before Regeneration)

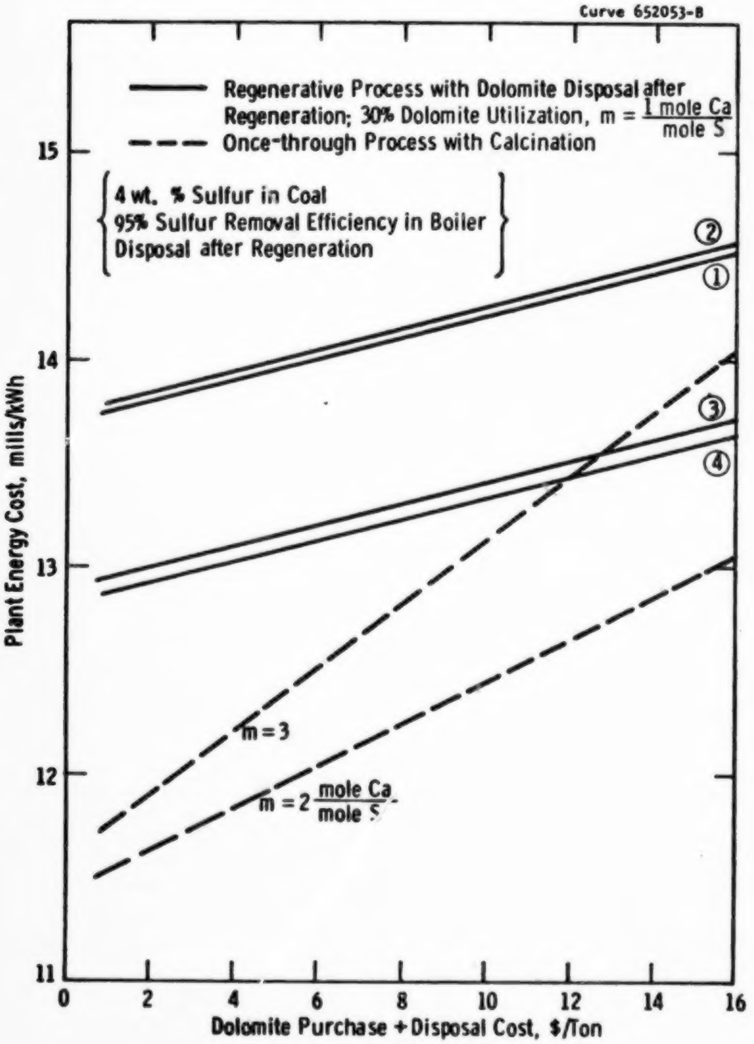

Figure 86: Plant Energy Cost (Disposal after Regeneration)

once-through operation. Further details of the once-through process are given in Appendix I and cost comparisons are considered in greater detail in Appendix K.

4.6 PROCESS PERFORMANCE

The fluidized bed plant heat rate, regeneration process temperature control, regeneration process turndown, and the overall environmental comparison of the regeneration processes are considered. These performance factors are necessary items to examine in order to understand the value of the various processes considered.

4.6.1 Plant Heat Rate

The plant heat rate is shown in Figure 87 for the four base design cases and for once-through operation. The heat rate is expressed as a function of the process sulfur load, W_S $(\epsilon - mx_S)$, for the regenerative processes or the factor W_S m x 10^{-1} for the once-through operation. For the once-through operation the resulting heat rate for the cases of calcination in the boiler and carbonation in the boiler are shown. Values of the regenerative process sulfur load may approach 0.06 for very high sulfur coals, while the factor $W_S m$ x 10^{-1} for once-through operation would probably never exceed 0.025 even with very high sulfur coals. Once-through operation or the 1 atmosphere one-step process yield the lowest plant heat rates. The 10 atmosphere one-step regeneration with 1% SO_2 in the regenerator product yields the highest plant heat rate.

4.6.2 Temperature Control

The feasibility of temperature control of the regeneration processes has been investigated and is discussed in Appendix G. The major consideration of temperature control feasibility is whether or not the required amount of energy can be supplied or withdrawn from the regenerator vessels while maintaining the specified temperature and concentration of SO_2 or H_2S in the regenerator product. It has already been stated that temperature control cannot be obtained in the

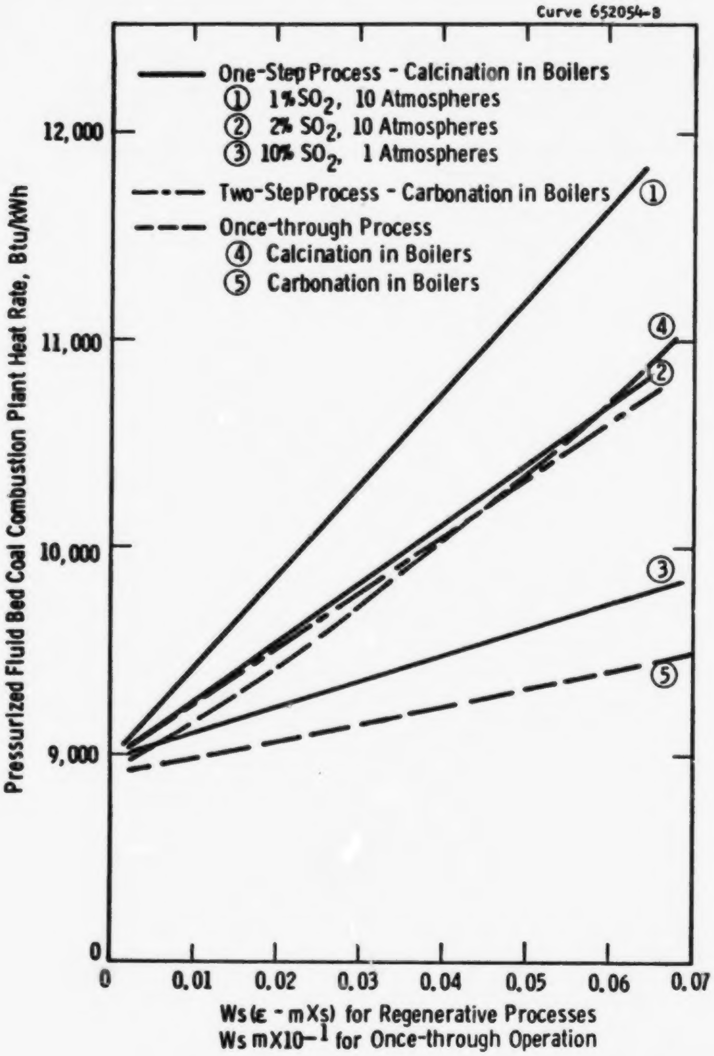

Figure 87: Plant Heat Rate with Regeneration Alternatives

regeneration processes if the sorbent is calcined in the boiler with the one-step process. This problem was handled by using separate vessels to process the sorbent before regeneration.

The two-step process is a low-temperature process compared to the one-step regeneration process, and temperature control is easily obtained in this process. With the one-step process, the temperature of the reducing gas entering the regenerator vessel becomes higher as the SO_2 mole fraction in the regenerator product increases. For the 10 atmosphere case where SO_2 mole fractions are 1 to 2%, the temperature control is feasible without excessively high temperatures; but with the 1 atmosphere case, where mole fractions up to 20% could be expected at 2000°F, temperature control is not possible. For this reason, the one-step process at 1 atmosphere pressure requires a reduced vessel temperature of 1900°F and preheating of the incoming dolomite to 1900°F in a separate vessel, yielding an SO_2 mole fraction of about 10%. The highest SO_2 mole fraction which would give feasible temperature control would be about 12 to 13%, without having heat transfer surface in the reactors.

4.6.3 Process Turndown

The base designs have assumed that four regenerator vessels are used by the regeneration processes (one for each boiler module) to minimize the turndown required for each vessel. Though a variety of turndown concepts are possible, a most attractive concept appears to be for the regenerator temperature and pressure to be maintained constant during turndown to yield a constant SO_2 or H_2S concentration to the sulfur recovery element during turndown. The sorbent circulation rate would be maintained constant during turndown to provide simpler control of the sorbent circulation equipment. The fluidization velocity in each vessel would be proportional to the fractional load on each vessel during turndown. With four regenerator modules, no more than 50% turndown of each vessel would be required. Turndown of other elements in the regeneration process have not been considered in detail, though the reducing-gas producer element and the sulfur recovery element may be problem areas, and multiple modules may be required to obtain high performance during turndown. The option of operating once-through during turndown at levels where the regeneration process performance is low should be considered. Some details of regenerator turndown are presented in Appendix G.

4.6.4 Process Environmental Comparison

The four base designs for regeneration are compared with once-through operation of the pressurized fluid bed combustion power plant on the grounds of environmental factors in Table 36. For each base design two columns are indicated: column A for disposal of dolomite before regeneration and column B for disposal following regeneration. For once-through operation two fresh dolomite feed rates are considered: m = 3.0 and m - 1.5 moles Ca/mole S fed which corresponds to 31.6 and 63.3% dolomite utilization in the boilers, respectively. The design basis is summarized on the table.

The first two factors, plant heat rate and plant energy cost, indicate the overall impact of the processes on the energy picture. The plant heat rate and the plant energy cost should be lower for once-through operation than for the regenerative processes, even with a high dolomite cost of $10/ton. Other factors of importance are the plant raw materials and the plant outputs. Coal and methane requirements are greater with the regenerative processes than they are with the once-through operation at the cost of the increased dolomite consumption required for once-through operation. Ash output is comparable to dolomite waste for the regenerative processes. Sulfur removal efficiency is expected to be greater for once-through operation than for regenerative operations, as indicated in the table, because of reduced chemical processing.

If the total United States electrical utility industry were to convert completely to pressurized fluid bed combustion, then the rate of coal consumption would be greater for conversion to regenerative operation

TABLE 36

COMPARISON OF PROCESSES - ENVIRONMENTAL IMPACTS[a,b]

Resources	One-Step-10 Atm				One-Step-1 Atm		Two-Step		Once-through	
	1% SO_2		2% SO_2		10% SO_2				m=3.0	m=1.5
	A	B	A	B	A	B	A	B	A	B
Plant heat rate, (Btu/kWh)[c]	10,300	10,720	9,830	10,100	9,350	9,470	9,780	10,000	9,250	9,080
Plant energy cost, (mills/kWh)	13.88	14.61	13.21	13.67	13.13	13.59	13.94	14.67	13.45	12.30
Raw materials[d]										
Total coal input, (tons/hr)	244	252	233	238	223	226	234	240	215	215
Dolomite input, (tons/day)	1,040	1,040	1,040	1,040	1,040	1,040	1,040	1,040	3,120	1,560
Methane input, (MM scf/day)	2.82	3.67	2.67	3.45	2.46	3.20	0	0	0	0
Plant Exports										
Waste dolomite, (tons/day)	890	770	890	770	890	770	1,080	1,070	2,700	1,640
Ash output, (tons/day)[e]	585	605	560	571	535	543	562	576	516	516
Sulfur output, (tons/day)	158	164	151	155	137	138	153	156	0	0
Sulfur removal efficiency,[f] %	90	90	90	90	85	85	90	90	94	94
Percent increase in U.S. coal production relative to once-through operation for total conversion to fluid bed combustion	13.3	17.3	8.1	10.5	3.8	4.9	90	11.7	0	0
Percent of U.S. limestone/dolomite production required for complete conversion to fluid bed combustion[g]	10	10	10	10	10	10	10	10	30	15

[a] Column A for disposal before regeneration, Column B for disposal after regeneration

[b] Basis -

430,000 lb/hr coal feed to boilers
Wt % sulfur in coal = 4.6
Boiler sulfur removal efficiency = 95%
Dolomite utilization in boiler = 30%
Dolomite utilization after regeneration = 10%
Regenerative make-up rate = 1 mole Ca/mole S
Dolomite cost = $10/ton
Coal at 45¢/MM Btu
Methane at 80¢/MM Btu
Sulfur recovery efficiency = 90%
Base design assumptions, restrictions, specifications, etc.
Once-through case assumes calcination in boilers

[c] Includes methane input.

[d] Other materials for regenerative processes include process water, chemicals, and catalysts.

[e] 10% ash in coal.

[f] Sulfur removal efficiency is lower for the one-step process at 1 atmosphere pressure because the process tail gas is incinerated rather than recycled to the boiler.

[g] Limestone/dolomite production taken from J. J. O'Donnell, A. G. Sliger, "Availability of Limestones and Dolomites," presented at the Second International Lime/Limestone Wet Scrubbing Syposium, New Orleans, Louisiana, November 8-12, 1971.

than for conversion to once-through operation. The relative rate of coal consumption for regenerative and once-through operations is shown in Table 36 and ranges from 4 to 18% greater for regeneration. Consumption of the valuable fuel methane is another factor to be considered. The price paid for this reduction in coal and methane consumption (in terms of environmental impact) is indicated in the last factor in the table. For complete conversion of the United States electric utility industry to pressurized fluid bed combustion,about 10% of the present United States limestone/dolomite production would be required for regenerative operation and 15 to 30% for once-through operation.

These preliminary considerations should add some perspective to the advantages and disadvantages of regenerative and once-through operations.

4.7 ASSESSMENT

4.7.1 Economic Factors

The plant energy cost for pressurized fluid bed combustion, utilizing dolomite regeneration and once-through operation, is compared to the plant energy cost for a conventional coal-fired power plant using limestone wet scrubbing for pollution control in Figure 88. Curves for the one-step process with 1% SO_2 (10 atmospheres) and 10% SO_2 (1 atmosphere) are shown. The costs for the one-step process with 2% SO_2 and the two-step process are not shown because they have costs nearly identical to the costs for the one-step with 10% SO_2 (1 atmosphere) and the one-step with 1% SO_2 (10 atmospheres), respectively. The cases of dolomite disposal before and after the regeneration are shown. Both limestone and dolomite costs have been taken as \$2/ton. The effect of dolomite cost is shown in Figures 85 and 86.

Figure 88 indicates that the energy cost of the conventional plant increases at a slower rate with increased coal sulfur content than does the cost of the fluid bed combustion plant. The cost of the

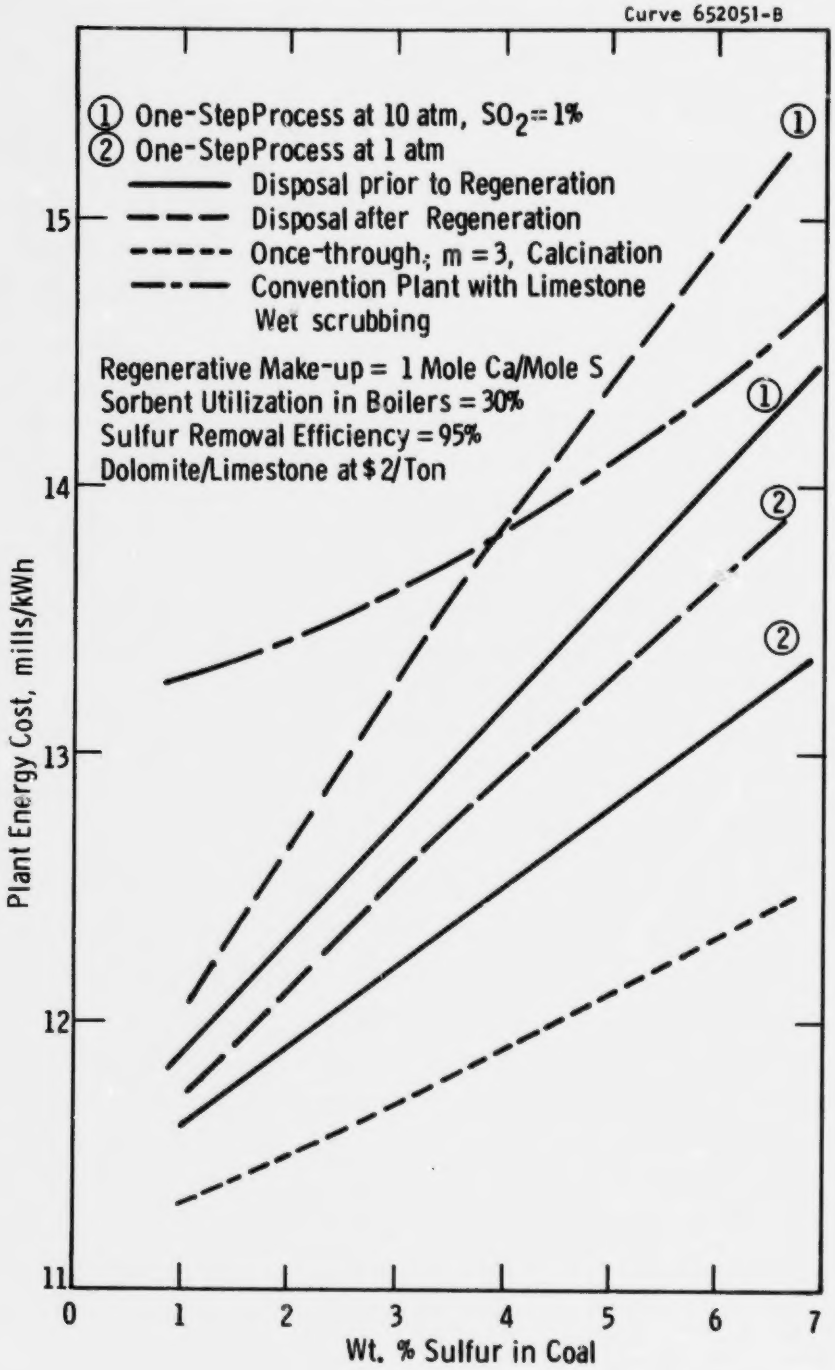

Figure 88: Comparison of Fluid Bed Power Plant with Conventional Power Plant

conventional plant is greater than the fluid bed combustion plant for all cases except the one-step process with 1% SO_2 with dolomite disposal following regeneration. The costs for the one-step process with 1% SO_2 with disposal before regeneration and the two-step process are comparable to the conventional plant energy cost at high-sulfur contents. The energy cost reduction with the one-step process with 10% SO_2 (1 atmosphere) and the one-step process with 2% SO_2 is indicated to be about 1 mill/kWh or greater. The cost reduction with once-through operation is shown to be about 2 mills/kWh with a low dolomite cost of $2/ton. The once-through cost reduction compared to the conventional plant will be about 1 mill/kWh with dolomite and limestone costs of $10/ton. The basis for conventional plant costs is discussed in Appendix N.

TABLE 37

GENERAL COMPARISON

- FBCOM with Regeneration

Optimistic Case	Pessimistic Case
Optimum boiler condition	Optimum boiler condition
Ca/S molar ratio, m = 1/2	Ca/S molar ratio, m = 1.5
Sorbent at $10/ton	Sorbent at $10/ton
Disposal before regeneration	Disposal after regeneration
No sulfur credit	No sulfur credit

- Once-Through

Optimistic Case	Pessimistic Case
Ca/S molar ratio, m = 1.2	Ca/S molar ratio, m = 3.0
Carbonation in boiler	Calcination in boiler
Sorbent at $10/ton	Sorbent at $10/ton

- Conventional Power Plant with Limestone Wet Scrubbing
- FBCOM Plant with Stack Gas Cleaning (Wellman-Lord)
- Conventional Power Plant with Wellman-Lord Stack Cleaning

A general comparison of energy costs between fluidized bed combustion with dolomite regeneration, fluidized bed combustion with once-through operation, fluidized bed combustion with Wellman-Lord stack gas cleaning, a conventional power plant with limestone wet scrubbing, and a conventional power plant with Wellman-Lord stack gas cleaning is shown in Figure 89. Table 37 indicates the conditions applied in Figure 89, with costs ranging between optimistic conditions and pessimistic conditions for regenerative and once-through fluid bed combustion power plants. Figure 90 indicates the effect of the plant capacity factor on the energy cost of regenerative and once-through fluid bed power plants, while Figure 91 shows the break-even cost of clean fuel as a function of the plant capacity factor. Evidently, the once-through operation becomes more economically attractive than the regenerative operation as the plant capacity factor decreases. The pressurized fluid bed combustion plant using the Wellman-Lord stack gas cleaning process has many potential improvements in efficiency, because no sorbent is present in the fluid bed boiler, which may reduce the energy cost below that indicated in Figure 91. The cost should be attractive compared to a conventional plant.

In general, the potential exists for pressurized fluid bed combustion power plants to reduce the cost of electrical energy generated by conventional power plants by 5 to 15% if the regeneration and once-through performance projected at the present time can be realized in commercial application.

4.7.2 Environmental Factors

Tables 38, 39, and 40 compare environmental factors for the alternative power plant concepts considered. Resources and wastes are listed in the tables. Table 38 compares regenerative fluid bed combustion and once-through operation. Table 39 compares once-through operation of the fluid bed combustion power plant to the fluid bed combustion plant with Wellman-Lord stack gas cleaning. Table 40 compares the fluid bed combustion plant with Wellman-Lord stack gas cleaning to a conventional coal-fired plant with Wellman-Lord stack gas cleaning.

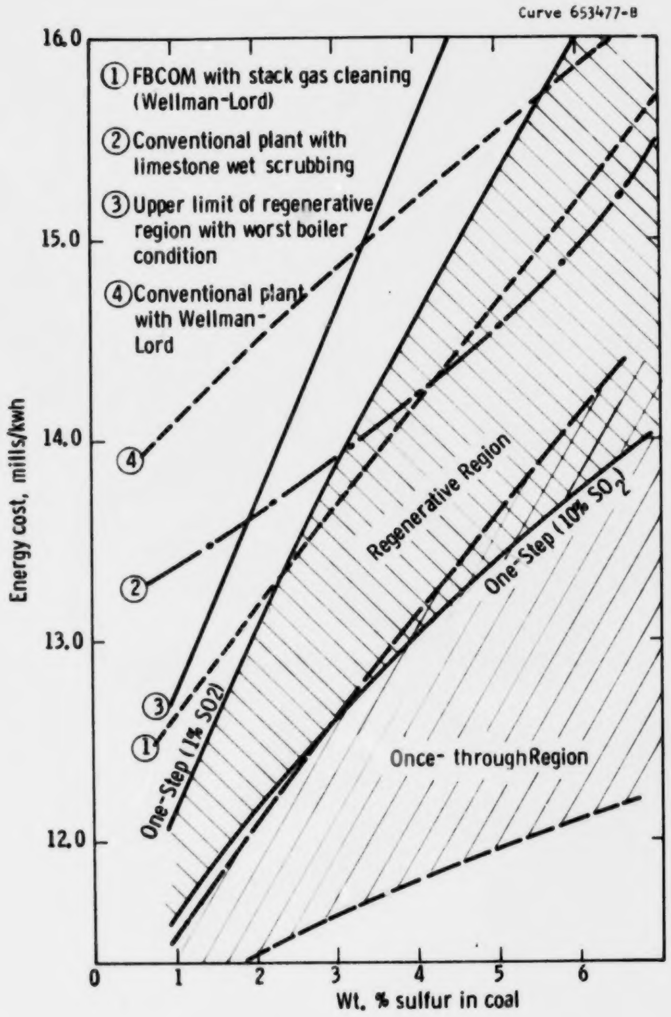

Figure 89: General Comparison of Energy Costs

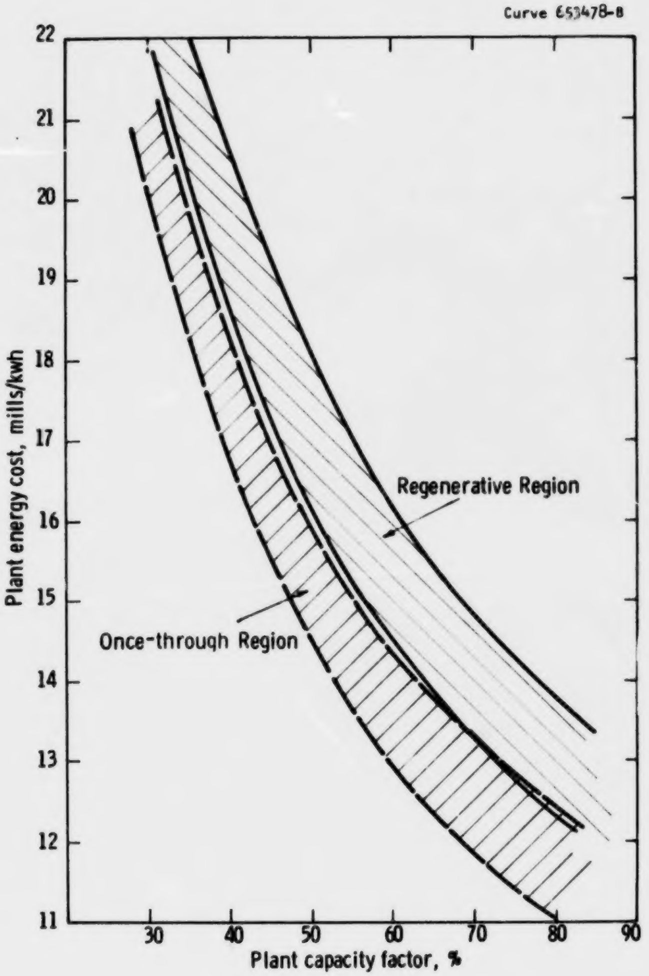

Figure 90: Effect of Capacity Factor

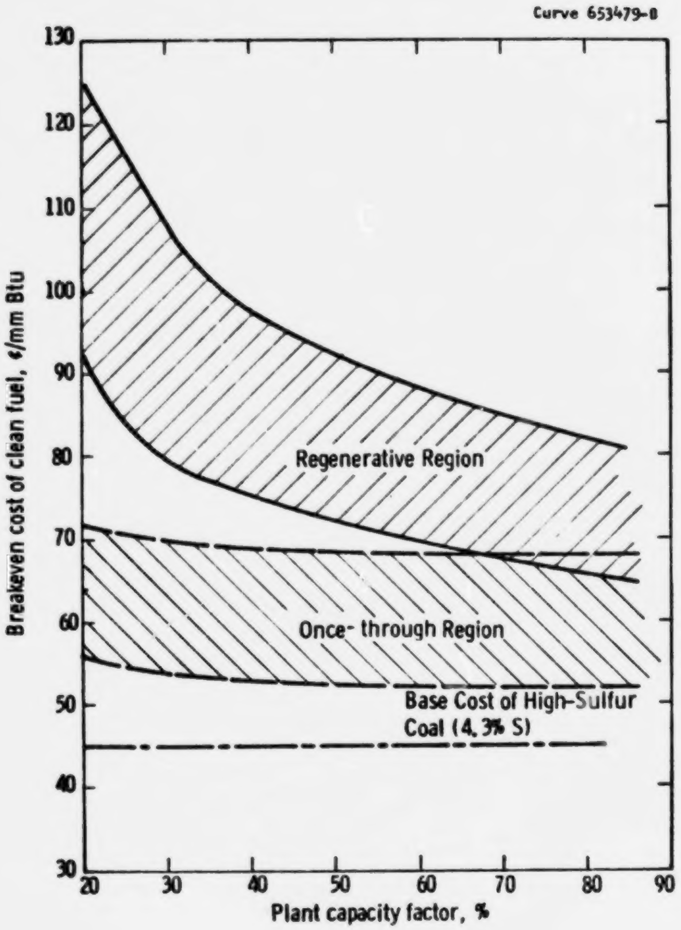

Figure 91: FBCOM with Low-Sulfur Coal

TABLE 38
COMPARISON OF REGENERATION WITH ONCE-THROUGH SYSTEM[a]

Item	One-Step		Two-Step	Once-through
	1 atm	10 atm (2% SO_2)		
Resources				
Plant heat rate, (Btu/kWh)	9,350	9,830	9,780	9,080
Coal (tons/hr)	223	233	234	215
Methane (MM scf/day)	2.46	2.67	0	0
Dolomite (tons/day)	1,040	1,040	1,040	1,560
Sulfur (tons/day)	145	151	153	0
Wastes				
Air pollution				
Overall SO_2 removal, %	90	90	90	94
NO_x, lb $NO_2/10^6$ Btu	0.1-0.2	0.1-0.2	0.1-0.2	0.1-0.2
Particulates, lb/10^6 Btu	0.05-0.15	0.05-0.15	0.05-0.15	0.05-0.15
Thermal, relative	1	1	1	1
Solids, tons/day				
dolomite	890	890	1,080	1,640
ash	535	560	562	516

[a] Basis: 635 MW(e), 4.6% Sulfur in Coal, 95% boiler sulfur removal

TABLE 39

COMPARISON OF ONCE-THROUGH WITH STACK GAS CLEANING SYSTEM[a]

Item	Once-Through Fluid Bed Combustion Plant m=1.55	Fluid Bed Combustion Plant with Wellman-Lord
Resources		
Plant heat rate, Btu/kWh	9,080	9,700
Coal, relative	1	1.08
NaOH, tons/day	0	20
Sulfur, tons/day	0	135
Methane, MM scf/day	0	2.45
Dolomite, tons/day	1,560	0
Wastes		
Air pollution		
Overall SO_2 removal efficiency, %	94	90
NO_x, lb $NO_2/10^6$ Btu	0.1-0.2	0.1-0.2
Particulates, $lb/10^6$ Btu	0.05-0.15	?
Thermal, relative	1	1.08
Liquid, tons/day (ft^3/day)	0	300 (10,000) - sodium sulfate, sodium thio sulfate, polythionates solution with some particulates
Solids, tons/day (ft^3/day)	1,640 (14,000) - dolomite 516 - ash	560 - ash

[a] Basis: 635 MW(e), 4.6% sulfur in coal, 95% boiler sulfur removal.

TABLE 40

COMPARISON OF STACK GAS CLEANING ON CONVENTIONAL AND FBB PLANTS[a]

Item	Fluid Bed Combustion Plant with Wellman-Lord	Conventional Power Plant with Wellman-Lord
Resources		
Plant heat rate, Btu/kWh	9,700	10,000
Coal, relative	1	1.03
NaOH, tons/day	20	20.6
Sulfur, tons/day	135	140
Methane, MM scf/day	2.45	2.55
Wastes		
Air pollution		
Overall SO_2 removal efficiency, %	90	90
NO_x, lb $NO_2/10^6$ Btu	0.1-0.2	0.05-0.15
Particulates, $lb/10^6$ Btu	?	?
Thermal, relative	1	1.03
Liquid, tons/day	300	309
Solids, tons/day	560-ash	577-ash

---

[a] Basis: 635 MW(e), 4.6% sulfur in coal, 95% boiler sulfur removal

Regenerative operation requires more coal and methane than once-through operation while once-through operation requires more dolomite disposal. The net environmental effect of these factors must be considered with great care. Once-through dolomite waste levels may be reduced to values nearly as low as with regenerative operation by obtaining high utilization. $CaSO_4$ is an inert material having many potential uses. Coal and methane are extremely valuable natural resources.

The Wellman-Lord process requires more coal and make-up NaOH than the once-through process. The liquid disposal stream indicated in Table 39 for the Wellman-Lord process can be eliminated by evaporation to dryness using a low-pressure stream. NaOH manufacture requires the expenditure of large amounts of energy and is equivalent to about a 2% efficiency reduction in the fluid bed combustion power plant using the Wellman-Lord process.

The fluidized bed combustion plant using the Wellman-Lord stack gas cleaning process compares favorably with a conventional coal-fired power plant using the Wellman-Lord process. Further improvement in the performance of the fluid bed combustion plant using the Wellman-Lord process is possible.

4.7.3 Base Design Feasibility

While both optimistic and pessimistic cost assumptions have been made throughout the base regeneration process designs, it has been optimistically assumed that all the process steps are commercially feasible. Some process steps are questionable, as noted in Table 41 where the major problems associated with each of the regeneration processes are considered. The questionable areas in the regeneration process are such as will lead to reduced power plant availability and high maintenance costs compared to the once-through operation. Higher energy costs than have been indicated for the process base designs may also result from the factors discussed in Table 41.

TABLE 41

ASSESSMENT OF BASE DESIGNS

LOW PRESSURE ONE-STEP REGENERATION PROCESS

Excessively High Reducing-Gas Inlet Temperature to Regenerator

- > 3000°F even with sorbent preheated to 1900°F for 10% SO_2
- Sorbent may be deactivated by high temperature near distributor - fusion of impurities, increased crystal size.
- Materials problem
- Inlet temperature can be reduced by decreasing the SO_2 fraction below 10% by (1) increasing the regenerator pressure or (2) reducing the regenerator temperature below 1900°F. Option (2) may lead to excessive CaS formation and possible eutectic formation with $CaSO_4$.
- Partial combustion of coal in the regenerator rather than externallv may eliminate the high-temperature problem but may introduce other problems.

Complex Sorbent Circulation Element

- Includes lock hoppers, valves and feeders, sorbent cooling and heating vessels
- Represents > 50% of process capital investment
- Involves potential sorbent losses, attrition, deactivation, and sulfur emissions
- Cost is sensitive to sorbent circulation rate; base design assumes change in sorbent utilization of 20% (optimistic assumption)
- Similar heating and cooling equipment has been operated commercially - moving bed shaft kiln at 1900°F with gas, oil, or coal as fuel; fluid bed aluminum hydroxide calciner at ∿ 2000°F; FluoSolids staged lime kiln by Dorr-Oliver; Pebble bed heat exchanger.
- Dense phase sorbent transport rather than dilute phase could eliminate lock hoppers, valves, feeders, and the necessity for cooling the sorbent. This would require an extremely high elevation of the regenerator and long stand legs.

TABLE 41 (Continued)

Commercial Feasibility of Reducing-Gas Generation from Coal

- Lurgi and Wellman-Galusha are commercial coal gasifiers for atmospheric pressure.
- May require size coal, steam; may be low efficiency with difficult control of outlet temperature and $CO + H_2$ concentration
- If coal can be partially combusted directly in the regenerator with the same performance assumed in the base design, the capital investment would be reduced by ∿ 10%, assuming the regenerator vessel cost does not increase (optimistic assumption).

Sorbent from Boiler in Carbonated Form

- Requires separate calcination system before regeneration
- Dolomite retains activity for CO_2 pick-up (∿ 60% still active after 10 cycles). Limestone is less active (∿ 33% after 10 cycles).
- Increases complexity and cost (∿ 60% increase in base design energy cost for 100% recarbonation) - sensitive to sorbent circulation rate
- Increased potential for sorbent losses, attrition, deactivation, and sulfur emissions
- Commerical calcining processes are well established.
- Also, conditions may change from calcining to carbonating in the boiler during turndown.

Sulfur Recovery

- Allied Chemical has commercial process for sulfur recovery down to 4 to 5% SO_2 (∿ 12 \$/kw for 10% SO_2).
- Shell report on sulfur recovery used in base design has optimistic costs (∿4 \$/kw for 10% SO_2).
- Contaminants in regenerator product gas may cause problems (O_2, H_2O, tars, metals, particulates).

TABLE 41 (Continued)

- Dilute or concentrated sulfuric acid may be recovered from dilute SO_2 gases - demonstrated by Parsons Co. on smelter's gas down to 2% SO_2.

Complex Process Control

- Integration of the regeneration process elements to follow the power plant load will be difficult. Each element has its own turndown problems.
- The constant load concept is a possibility with this process. Once-through operation during stages of turndown is a possibility.

HIGH-PRESSURE ONE-STEP REGENERATION PROCESS

Commercial Feasibility of Sulfur Recovery

- Base design assumes highly optimistic costs in Shell report. (3 to 4 \$/kw not including incineration at 1 to 2% SO_2).
- Allied Chemical process would cost ∿ 25 \$/kw (0.95 mills/kWh), for 3% SO_2 and probably > 50 \$/kw (1.8 mills/kWh) for 1% SO_2. Allied Chemical process implies greater complexity than Shell design.
- Contaminants such as tars and particulates have been neglected.
- Sulfuric acid recovery is probably not commercial for < 2% SO_2.

High Reducing-Gas Inlet Temperature to Regenerator

- > 2300°F for 1% SO_2 and > 2500°F for 2% SO_2. Increasing during turndown - > 2800°F at 50% turndown and 2% SO_2.
- Possible sorbent deactivation near distribution
- If coal can be partially combusted directly in the regenerator, with no change in performance and no increase in the cost of the regenerator vessel, then the capital investment can be reduced by about 17% but with no fuel saving. About ∿ 10% reduction in regeneration process energy cost

TABLE 41 (Continued)

or ∿ 2% reduction in power plant energy cost. May eliminate high-temperature problem - may introduce other problems.

Reducing Gas Generation from Coal

- Base design assumes a very low level of $CO + H_2$ required in the reducing gas (1.5% $CO + H_2$ for 1% SO_2 and 3% $CO + H_2$ for 2% SO_2); it has not been experimentally demonstrated.
- With low $CO + H_2$ assumption the reducing gas can be generated using a coal combustor operated slightly below stoichiometric combustion rather than coal gasifier.
- Commercial coal combustors (slagging) are available, but Lurgi is the only commercial gasifier at pressure.
- If higher $CO + H_2$ levels must be used, then the coal consumption and cost for reducing-gas production increases ∿ linearly with the fraction $CO + H_2$. CaS production also increased. (Doubling the base design fractions of $CO + H_2$ increased the process energy cost by ∿ 30%).

Sorbent from Boiler in Carbonated Form

- Requires separate calcination system before regeneration
- Dolomite retains activity for CO_2 pick-up. Limestone is less active.
- Increased complexity and cost-sensitive to sorbent circulation rate
- Increased potential for sorbent losses, attrition, deactivation, and sulfur emissions
- Commercial calcining processes are well-established
- Conditions may change from calcining to carbonating in the boiler during turndown

Complex Process Control

- Integration of the regeneration process elements to follow the power plant load will be difficult. Each element has its own turndown problems.

TABLE 41 (Continued)

TWO-STEP REGENERATION PROCESS

Commercial Feasibility of Reducing Gas Generation from Coal

- Lurgi is the only commercial coal gasifier at pressure.
- May require sized coal and steam; may be low efficiency with difficult control of outlet temperature and CO + H_2 concentration.

Commercial Feasibility of CO_2 Recovery

- Commercial processes for CO_2 recovery exist (such as the Hot Carbonate process).
- The effects of contaminants which may be present in Claus plant tail gas or power plant stack gas and the turndown characteristics have not been examined.

High CaS Content in Regenerated Sorbent

- $CaSO_4$ reduction step gives high conversion to CaS in laboratory studies. H_2S generating step gives poor conversion to $CaCO_3$ in batch tests.
- High CaS content in regenerated sorbent may cause eutectic melt to form in boiler.
- Large circulation rates will be required with utilization changes of 5 to 10%. High utilization in the boiler will require a large quantity of reducing gas.

Sorbent in Calcined Form in Boiler

- Requires separate carbonation system before regeneration
- Increases complexity and cost of process. Cost increase is highly sensitive to the sorbent circulation rate.

TABLE 41 (Continued)

- Increased potential for sorbent losses, attrition, deactivation, and sulfur emissions. Disposal product may contain CaS.
- Conditions in boiler (calcination or carbonation) may change during turndown.

Sulfur Recovery

- Recovery of sulfur from low H_2S gas ($\geq$ 3%) streams is probably commercial, but also probably more costly than indicated in the Shell report used in the base design.

Complex Process Control

- Integration of the regeneration process elements to follow the power plant load will be difficult. Each element has its own turndown problems.

4.8 COMBUSTION OF LOW-SULFUR COALS

Considerations have so far been with the fluid bed combustion power plant operated regeneratively or once-through with coals having sulfur contents greater than 1 wt % and requiring high sulfur removal (∿ 95%). The case of low-sulfur coals which require little or no desulfurization is also of importance since such coals are attractive fuels and are still found in some areas of the United States.

The capital costs of the pressurized fluid bed combustion plant and the conventional power plant with limestone wet scrubbing are shown in Table 42. A 1 wt % sulfur coal is considered with 70, 50, and 0% desulfurization required. The plant energy cost is shown in Table 43 for the same cases. As was indicated in Figure 89, the cost differential between fluidized bed combustion and conventional power generation becomes larger as the coal sulfur content is reduced. An 11% reduction in energy cost is projected for the case of no desulfurization. With no desulfurization required, many improvements in the design of the fluidized bed combustion processes are possible. These are discussed in Appendix M.

TABLE 42

LOW-SULFUR COAL COMBUSTION - INVESTMENT ($/kw)

% Desulfurization of 1% Sulfur Coal	Fluid Bed Combustion		Conventional Plant with Limestone Wet Scrubbing
	Once-through	Regenerative[a]	
70	265.6	278.5	331.4
50	265.0	272.5	330.5
0	260.5		316.3

---

[a] Based on the one-step regeneration process at 10 atmospheres with 1% SO_2.

TABLE 43

LOW-SULFUR COAL COMBUSTION - ENERGY COST (mills/kWh)

% Desulfurization of 1% Sulfur Coal	Fluid Bed Combustion		Conventional Plant with Limestone Wet Scrubbing
	Once-through	Regeneration[a]	
70	11.31	11.86	13.36
50	11.24	11.58	13.29
0	11.12		12.51

---

[a] Based on the one-step regeneration process at 10 atmospheres with 1% SO_2.

4.9 CONCLUSIONS

- Pressurized fluid bed combustion combined cycle power generation plants with limestone/dolomite sulfur removal have lower energy costs and offer environmental advantages over conventional plants with stack gas cleaning.
- A once-through sulfur sorbent process is the most attractive -- economically and environmentally -- for first-generation pressurized fluid bed boiler combined cycle power plants. Energy costs could be reduced more than 10%, air emission criteria met, and the spent sorbent utilized or disposed of without further processing and with no environmental pollution.
- Once-through dolomite/limestone operation should be demonstrated on the pressurized pilot plants now being built* and the pressurized fluid bed combustion boiler demonstration plant.
- Further development of regeneration processes is required before they can be demonstrated on a commercial scale.
- Experimental work should continue to
  - Achieve high stone utilization for once-through operation
  - Improve present regeneration process concepts
  - Develop new regeneration processes
  - Study alternative sorbents to limestones/dolomites
- The cost advantage for a pressurized fluidized bed combustion plant over conventional plants with stack gas cleaning is achieved with high- or low- (e.g. 1%) sulfur content in the fuel.

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*EPA contracts to Esso and Argonne National Laboratory.

- The cost advantage for pressurized fluidized bed combustion plants over conventional plants may be greater using low-sulfur fuels not requiring sulfur removal.

5. PLANT OPERATION AND CONTROL

5.1 INTRODUCTION

A philosophy of operation and recommendations for controlling the pressurized fluid bed boiler power plant have been developed. In addition, an assessment has been made of the ability of the concept to meet utility operating requirements. This study presents a definition of procedures with estimates of key parameters.

It is assumed that the reader is familiar with the operation and control of large once-through boiler power plants and fluidized bed combustion boilers. Many normal power plant operating procedures and control techniques are referred to but not detailed or defined. The operating procedures contained in this report are based on the normal operation of conventional boiler and steam-turbine power plants. The control system recommendations are based on a large, hybrid control system utilizing digital computers and analog control capability.

5.1.1 Plant Criteria

A power plant that is to operate as a swing plant must be capable of starting up, changing load, and shutting down easily. In the plant design, the equipment configuration should be selected to provide considerable flexibility for plant operation.

Plant operation is based on the criteria that the plant

- Operate as a swing or base load plant
- Be capable of operation from 12-1/2% to 100% of rated capacity, at design conditions of 2400 psi and 1000°F superheat and 1000°F reheat
- Be capable of changing load at the rate of 5% per minute
- Operate with some of the plant components out of service for maintenance

- Be capable of starting up or shutting down components in several different sequences to ensure maximum usage of the plant equipment.

5.1.2 Control System Requirements

The recommended control system is intended to provide the total plant automation and control required to operate the plant efficiently, safely, and with the maximum amount of plant availability. To accomplish this, the following features must be provided:

- The plant automation required to start up and with the various plant components required for load changes
- A coordinated control system to operate all of the plant components properly
- An interface system so the plant can be operated remotely by the load dispatcher
- A loading scheme that will produce the best heat rate at all times
- An emergency loading scheme to load the plant in the minimum amount of time
- A predictive loading program to determine the plant configuration and loading sequence if the desired load and time requirements are known in advance
- An emergency or safety system to handle contingency conditions.

Since the operation of a plant must be understood before a control system can be recommended, the first portion of this report deals with the plant operation and the later portion with control system recommendations.

5.1.3 Pressurized Fluid Bed Boiler Power Plant Concept

Details of fluidized bed combustion boilers and the base 635 MW pressurized power plant design are presented elsewhere.[1] A review of the concept follows:

In the fluidized bed combustion boiler, sufficient air is added to the fuel to virtually complete the transformation of its carbon to CO_2 and its hydrogen to H_2O:

$$C + O_2 \rightarrow CO_2$$
$$H_2 + 1/2\ O_2 \rightarrow H_2O \quad .$$

Both these reactions produce large quantities of heat which are absorbed by generating steam.

Sulfur in the fuel is converted primarily to SO_2. A limestone or dolomite sorbent is utilized to remove this pollutant from the combustion gases:

$$\begin{pmatrix} CaCO_3 \\ CaO \end{pmatrix} + SO_2 + 1/2\ O_2 \rightarrow CaSO_4 + \begin{pmatrix} CO_2 \\ - \end{pmatrix} \quad .$$

The $CaSO_4$ can either be disposed of as a solid -- once-through operation -- or be processed to regenerate $CaCO_3$ (or CaO) sorbent and to recover sulfur as a useful product -- elemental sulfur, S, or sulfuric acid, H_2SO_4. Once-through operation could be advantageous if high utilization of the sorbent can be realized in the boiler. This would provide a simpler plant design and operation. Two processes appear most attractive for regenerating the stone:

- Reduction of calcium sulfate to calcium sulfide and subsequent regeneration with steam and CO_2:

$$CaSO_4 + \begin{pmatrix} 4\ H_2 \\ 4\ CO \end{pmatrix} \rightarrow CaS + \begin{pmatrix} 4\ H_2O \\ 4\ CO_2 \end{pmatrix}$$
$$CaS + H_2O + CO_2 \rightarrow CaCO_3 + H_2S$$

- Direct reduction of calcium sulfate to calcium oxide and SO_2 employing partially oxidized fuel as the reducing agent:

$$CaSO_4 + \begin{pmatrix} H_2 \\ CO \end{pmatrix} \rightarrow CaO + SO_2 + \begin{pmatrix} H_2O \\ CO_2 \end{pmatrix}$$

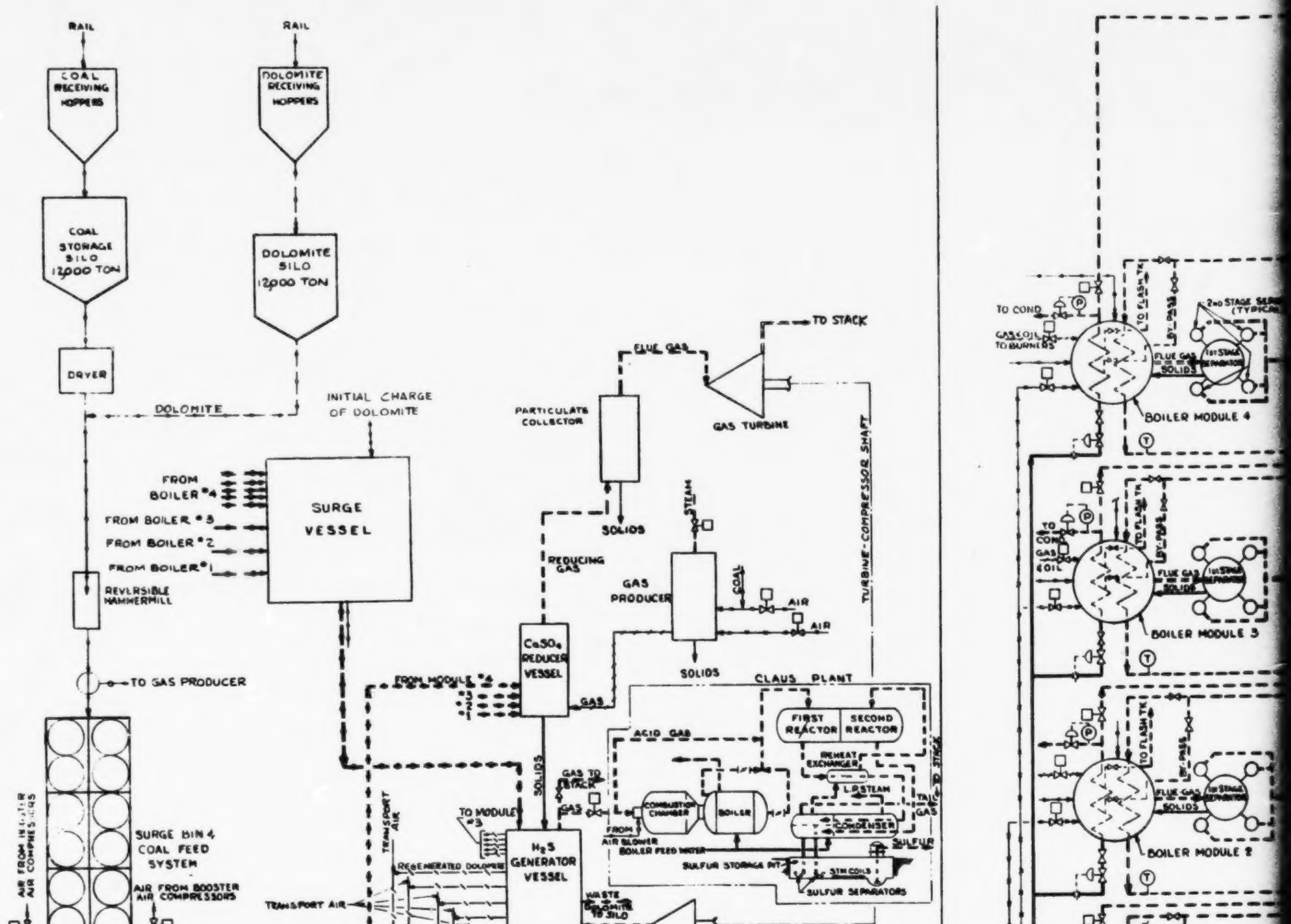
These alternatives are currently being studied to determine which should be used for commercial plants. The present evaluation assumes the first regeneration reaction system is used.

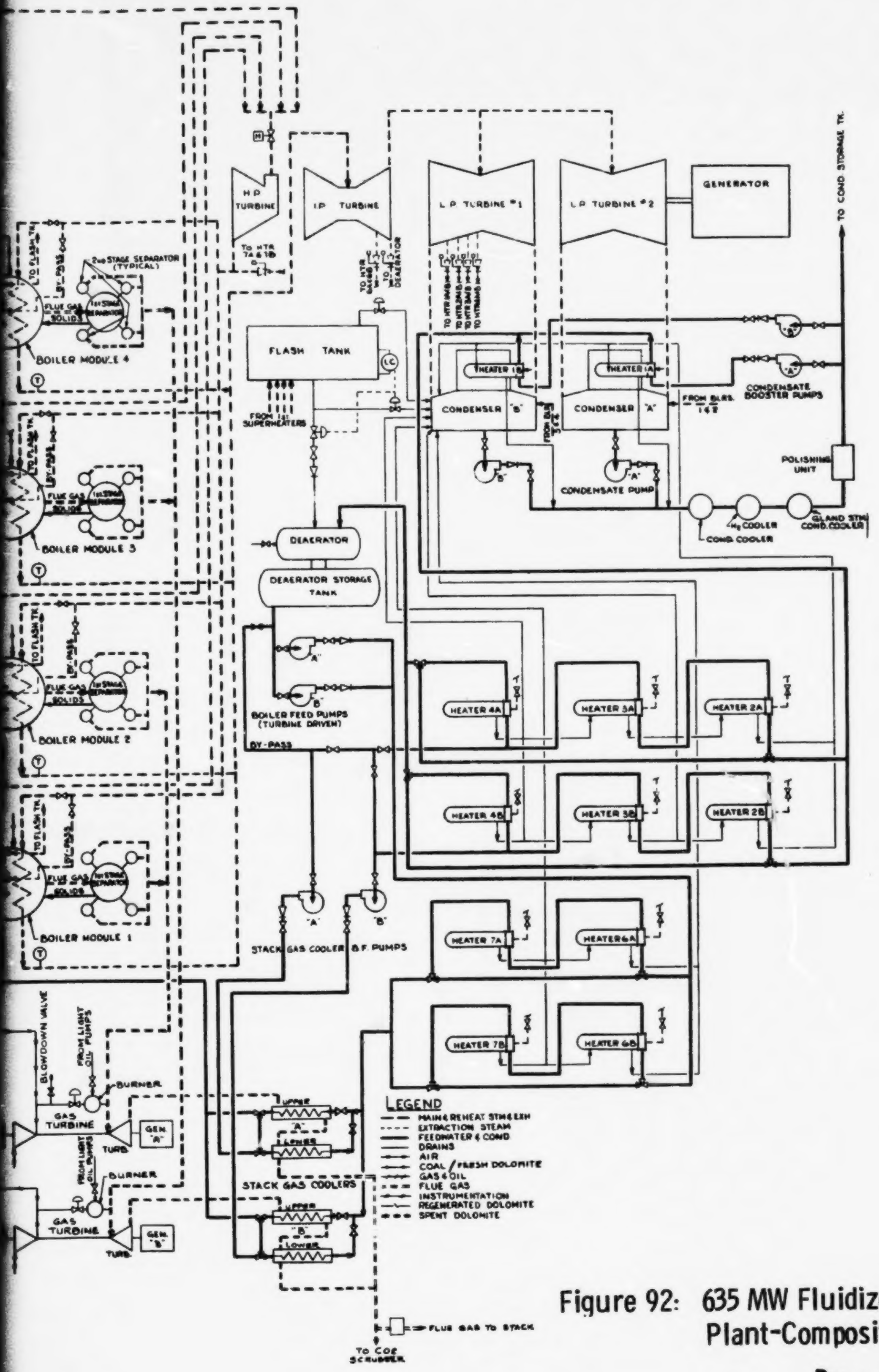
A fluidized bed combustion boiler is formed by an enclosure usually consisting of water walls -- abutting boiler tubes in which water and steam flow. Coal and sorbent are fed to the bed by pneumatic feeders extending through the air distributor or the water walls. From 60 to 70% of the heat released in burning the fuel with air is transferred to the water/steam in the tubes surrounding and submerged in the bed. Various tube configurations have been proposed -- horizontal tubes extending through the bed, horizontal tubes with serpentine bends in vertical platens or planes, and vertical tube walls passing up through the distributor plate, slicing the bed into narrow segments.

Operating at elevated pressure, a fluidized combustor requires a compressor to pressurize the air and to overcome the pressure loss over the fluidized bed combustor. The power cycle schematic utilizing coal (or oil) fluidized bed combustion at elevated pressure is illustrated in Figure 6, p. 19. At an operating pressure of 10 to 15 atmospheres, excess air of 10 to 15%, and a fluidizing velocity of 8 to 12 ft/sec, a depth of 8 to 15 feet is required to accommodate the heat transfer surface in the bed. Energy is recovered from the high-temperature gases by passing them directly into a gas turbine and expanding them to atmospheric pressure. This expansion lowers the temperature of the gases by 600 to 800°F, thereby reducing the amount of surface required to recover heat from the combustion gases leaving the fluidized bed. The pressurized system can be operated at higher excess air ratios. This increases the fraction of gas-turbine power, reduces combustible losses from the boiler, and increases the waste heat recovery after the gas turbine. This may allow for improved plant efficiencies and provide an effective means for achieving continuous, rapid plant turndown.

The boiler design consists of four modules. The modularized design provides for a maximum of shop fabrication and standardization and assists in meeting the turndown requirements for the plant. Each module includes four primary fluidized bed combustors, each in turn containing a separate boiler function -- one bed for the preevaporator, two beds for the superheater, and one bed for the reheater. Evaporation takes place in the water walls. All of the boiler heat transfer surface is immersed in the beds, except for the water walls. There is no convection heat transfer surface since the maximum allowable bed temperature is less than the state-of-the-art gas-turbine temperature. The fluidized bed combustors are stacked because of the inherent advantages in vertical gas circuitry, steam circuitry, and pressure vessel design. Each module contains a separate fluidized bed or carbon burn-up cell to complete the combustion of any carbon elutriated from the primary beds. A separate bed may not be required since carbon losses may be low enough in the proposed pressurized boiler design with deep beds. A carbon burn-up cell (CBC) is definitely not envisaged if the system is operated with high excess air. This also has the advantage of increasing plant performance by increasing the fraction of gas-turbine power. A simplified drawing of a module is shown in Figure 2, p. 3. The 318 MW plant module can be shop fabricated and shipped by rail. The 635 MW plant module is designed to be shipped in sections, each shop-fabricated. The primary beds for a 318 MW plant are approximately 5 ft x 7 ft. The bed depths are approximately 12 feet -- sufficient for the required heat transfer surface. The carbon burn-up cell (CBC) is approximately 2 ft x 7 ft. The CBC contains no submerged surface in the bed. The submerged tube bundles are formed by vertical tube platens or planes, each platen a continuous boiler tube in a serpentine arrangement. The heat transfer surface can be viewed as horizontal tubes. The preevaporator and reheater contain 1-1/2-in. diameter tubes; the evaporator water walls and reheater bed contain 2-in. diameter tubes.

A preliminary design plant layout, operating philosophy, and a cost estimate were prepared for the 635 MW plant.[1] A composite flow diagram for the base design is shown in Figure 92. A summary of the boiler design and the plant layout has been presented.[2] A summary of part load operation is presented in Appendix O.


RAIL
COAL RECEIVING HOPPERS
RAIL
DOLOMITE RECEIVING HOPPERS
COAL STORAGE SILO 12,000 TON
DOLOMITE SILO 12,000 TON
DRYER
DOLOMITE
INITIAL CHARGE OF DOLOMITE
FROM BOILER #4
FROM BOILER #3
FROM BOILER #2
FROM BOILER #1
SURGE VESSEL
REVERSIBLE HAMMERMILL
TO GAS PRODUCER
SURGE BIN & COAL FEED SYSTEM
AIR FROM BOOSTER AIR COMPRESSORS
COAL
FLUE GAS
TO STACK
GAS TURBINE
PARTICULATE COLLECTOR
SOLIDS
STEAM
REDUCING GAS
GAS PRODUCER
COAL
AIR
AIR
TURBINE-COMPRESSOR SHAFT
CaSO₄ REDUCER VESSEL
FROM MODULE #4
GAS
SOLIDS
CLAUS PLANT
FIRST REACTOR
SECOND REACTOR
ACID GAS
REHEAT EXCHANGER
L.P. STEAM
TO STACK
TAIL GAS
COMBUSTION CHAMBER
BOILER
CONDENSER
FROM AIR BLOWER
BOILER FEED WATER
SULFUR
SULFUR STORAGE PIT
STM COILS
SULFUR SEPARATORS
GAS TO STACK
GAS
TO MODULE #3
SOLIDS
TRANSPORT AIR
REGENERATED DOLOMITE
H₂S GENERATOR VESSEL
WASTE DOLOMITE TO SILO
TRANSPORT AIR
COMPRESSOR
TO MODULE #2
TO MODULE #1
HEAT EXCHANGER
CONDENSED LIQUID
HOT REHEAT TO LP TURBINE
TO SURGE VESSEL
AIR
REHEAT BED
TO HP TURB.
COLD REHEAT FROM HP TURB.
TO SURGE VESSEL
2nd SH BED
TO FLASH TANK
TO FLASH TANK
AIR
TO SURGE VESSEL
1st SH BED
SPENT DOLOMITE
TRANSPORT AIR
CO₂ SCRUBBER SYSTEM
FLUE GAS FROM STACK GAS COOLERS
HOLDING VESSEL One(1) Per. BOILER MODULE
AIR
TO SURGE VESSEL
EVAP BED & WALLS
AIR
CARBON BURN-UP CELL
AIR
FEEDWTR
TO SURGE VESSEL
SOLIDS FROM 1st STAGE SEPARATOR
MODULE #4
LIGHT OIL TANK
LIGHT OIL PUMPS
TO MODULE 1, 2, & 3
TO GAS TURBINES
TO COND
GAS & OIL TO BURNERS
TO FLASH TK
BY-PASS
2nd STAGE SEPARATOR (TYPICAL)
1st STAGE SEPARATOR
FLUE GAS
SOLIDS
BOILER MODULE 4
BOILER MODULE 3
BOILER MODULE 2
BOILER MODULE 1
AIR TO PRIMARY INJECTORS
BOOSTER AIR COMPRESSOR
BLOWDOWN VALVE
FROM LIGHT OIL PUMPS
BURNER
GAS TURBINE
MOTOR "A"
TURB.
AIR TO PRIMARY INJECTORS
BOOSTER AIR COMPRESSOR
MOTOR "B"
GAS TURBINE
TURB.


H P TURBINE
I P TURBINE
L P TURBINE #1
L P TURBINE #2
GENERATOR
TO COND STORAGE TK.
TO HTR 7A & 7B
TO DEAERATOR
FLASH TANK
LC
FROM 1st SUPERHEATERS
HEATER 1B
HEATER 1A
CONDENSER "B"
CONDENSER "A"
FROM BLRS. 1 & 2
FROM BLRS. 3 & 4
CONDENSATE BOOSTER PUMPS
POLISHING UNIT
CONDENSATE PUMP
COND. COOLER
H2 COOLER
GLAND STM. COND. COOLER
DEAERATOR
DEAERATOR STORAGE TANK
BOILER FEED PUMPS (TURBINE DRIVEN)
BY-PASS
HEATER 4A
HEATER 3A
HEATER 2A
HEATER 4B
HEATER 3B
HEATER 2B
HEATER 7A
HEATER 6A
HEATER 7B
HEATER 6B
STACK GAS COOLER B.F. PUMPS
2nd STAGE SEPARATOR (TYPICAL)
1st STAGE SEPARATOR
TO FLASH TK.
FLUE GAS
SOLIDS
BOILER MODULE 4
BOILER MODULE 3
BOILER MODULE 2
BOILER MODULE 1
BLOWDOWN VALVE
FROM LIGHT OIL PUMPS
BURNER
GAS TURBINE
TURB.
GEN. "A"
GEN. "B"
UPPER
LOWER
STACK GAS COOLERS
FLUE GAS TO STACK
TO COE SCRUBBER
LEGEND
MAIN & REHEAT STM & EXH
EXTRACTION STEAM
FEEDWATER & COND
DRAINS
AIR
COAL / FRESH DOLOMITE
GAS & OIL
FLUE GAS
INSTRUMENTATION
REGENERATED DOLOMITE
SPENT DOLOMITE

635 MW FLUIDIZED BED
BOILER COMBINED CYCLE PLANT
COMPOSITE FLOW DIAGRAM

united engineers

Figure 92: 635 MW Fluidized Bed Boiler Combined Cycle Plant-Composite Flow Diagram

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5.2 UNIT START-UP

5.2.1 Introduction

Plant start-up, as well as large-scale load increases, requires placing different parts of the total plant in operation. The normal start-up of the plant is sequential; that is, the plant components are placed in operation in a definite order as follows:

- Gas turbine
- Boiler module
- Steam turbine
- Boiler module
- Gas turbine
- Boiler module
- Boiler module

This sequence consists of three different start-up procedures:

- Gas turbine, first boiler module, and steam turbine
- The second and fourth boiler module start-up. These two are the same.
- The second gas turbine and third boiler module.

Figure 93, steam and water cycle, and Figure 94, chemical recovery plant, are process schematics showing the control valves referred to in this section and their functional location in the plant.

Figures 95A, 95B, 96A, and 96B show in a simplified way the steps required to start up the first and second boiler modules.

The start-up procedures that follow are based on sequentially starting the entire plant from a cold condition.

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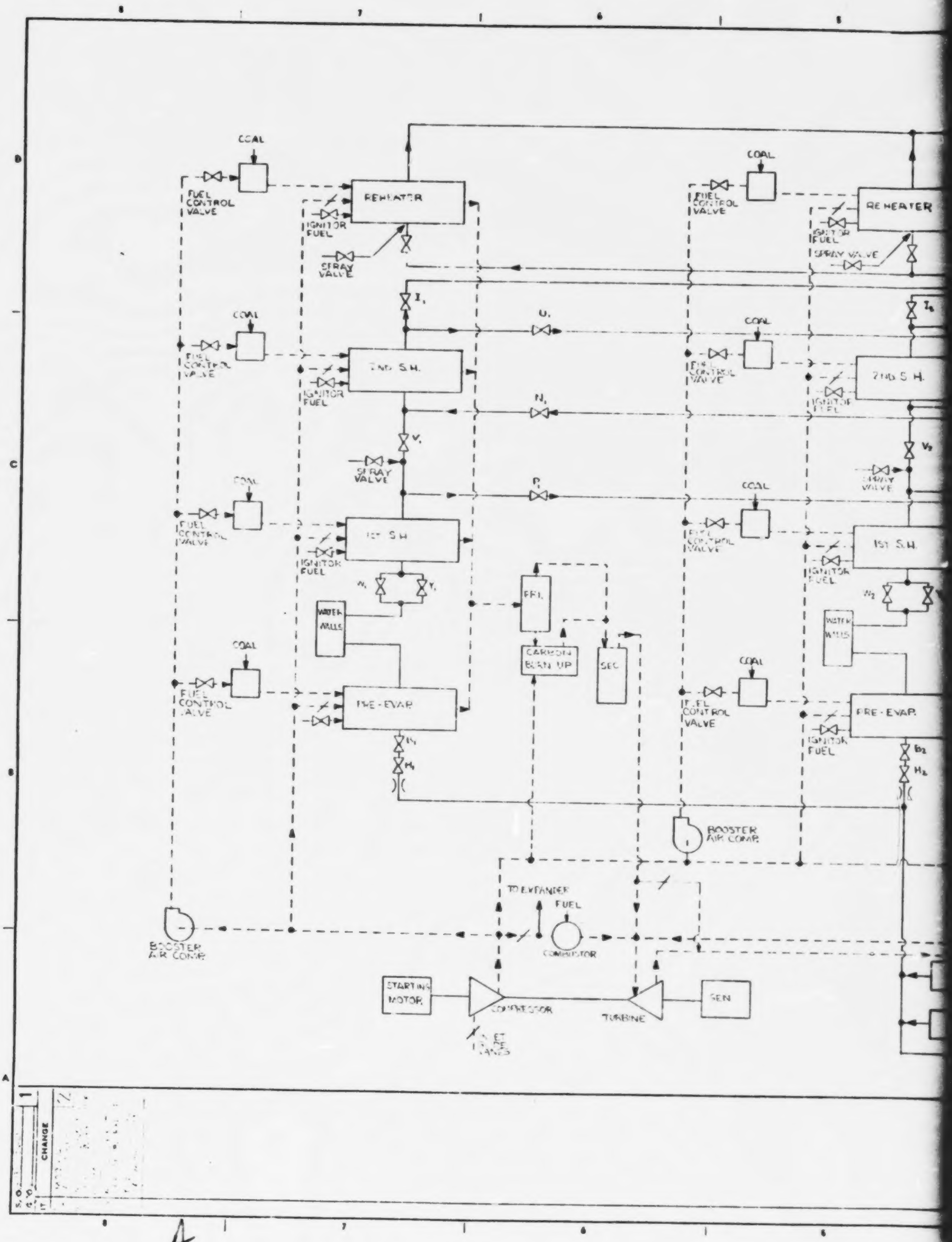

COAL
FUEL CONTROL VALVE
REHEATER
IGNITOR FUEL
SPRAY VALVE
I_1
U_1
2ND. S.H.
N_1
V_1
P_1
1ST. S.H.
W_1
Y_1
WATER WALLS
PFL
CARBON BURN UP
SEC
PRE-EVAP.
B_1
H_1
BOOSTER AIR COMP.
TO EXPANDER
FUEL
COMBUSTOR
STARTING MOTOR
COMPRESSOR
TURBINE
GEN
INLET GUIDE VANES
REHEATER
2ND. S.H.
I_2
V_2
1ST. S.H.
W_2
WATER WALLS
PRE-EVAP.
B_2
H_2
CHANGE
D
C
B
A
8
7
6
5

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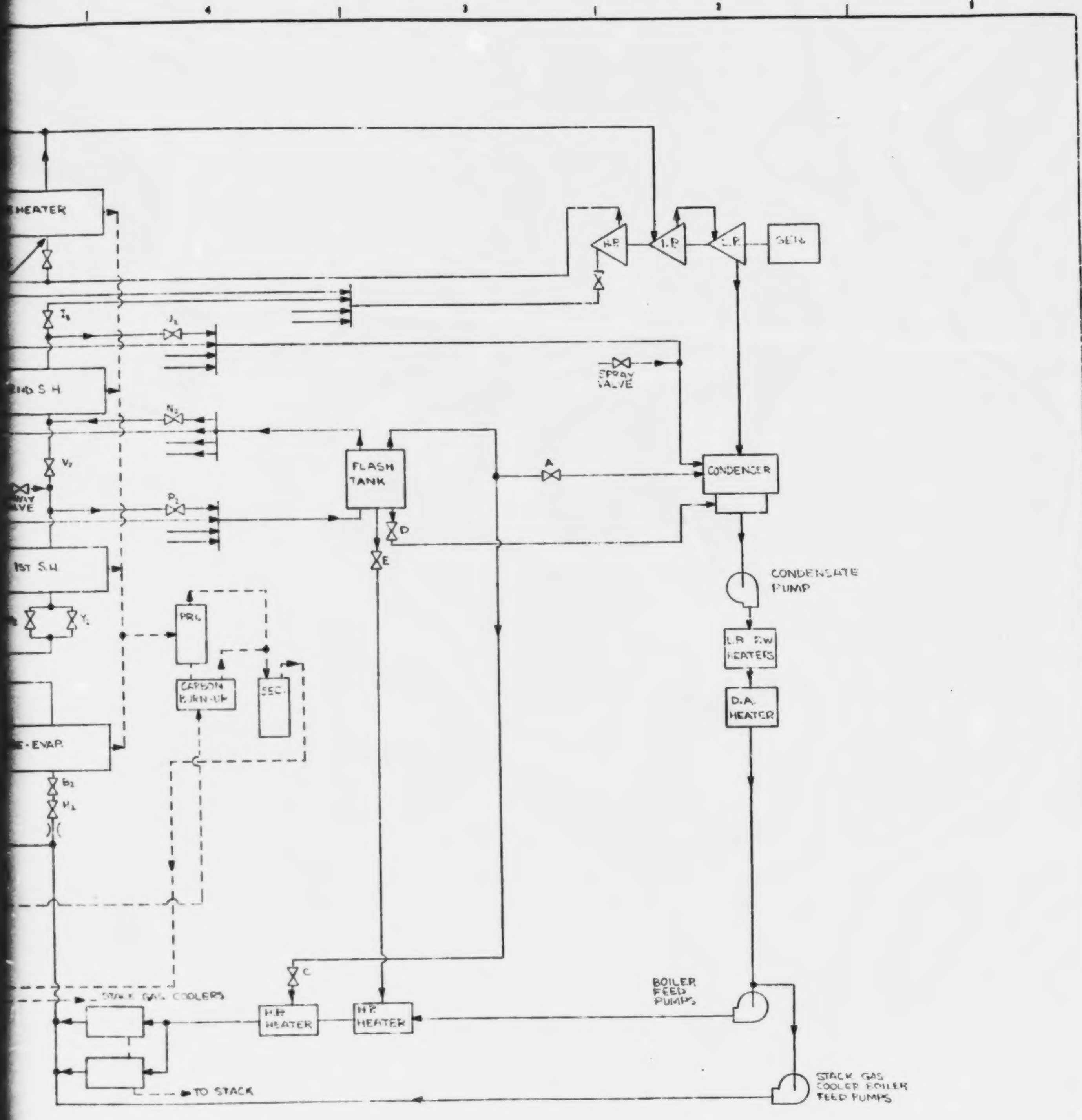
4
3
2
1
HEATER
2ND S.H
1ST S.H
E-EVAP.
H.P.
I.P.
L.P.
GEN.
J_2
N_2
P_2
V_2
Y_1
B_2
H_2
SPRAY VALVE
FLASH TANK
A
D
E
CONDENSER
CONDENSATE PUMP
L.P. F.W HEATERS
D.A. HEATER
PRI.
CARBON BURN-UP
SEC.
C
BOILER FEED PUMPS
STACK GAS COOLERS
H.P. HEATER
H.P. HEATER
TO STACK
STACK GAS COOLER BOILER FEED PUMPS

Figure 93: Steam and Water Cycle

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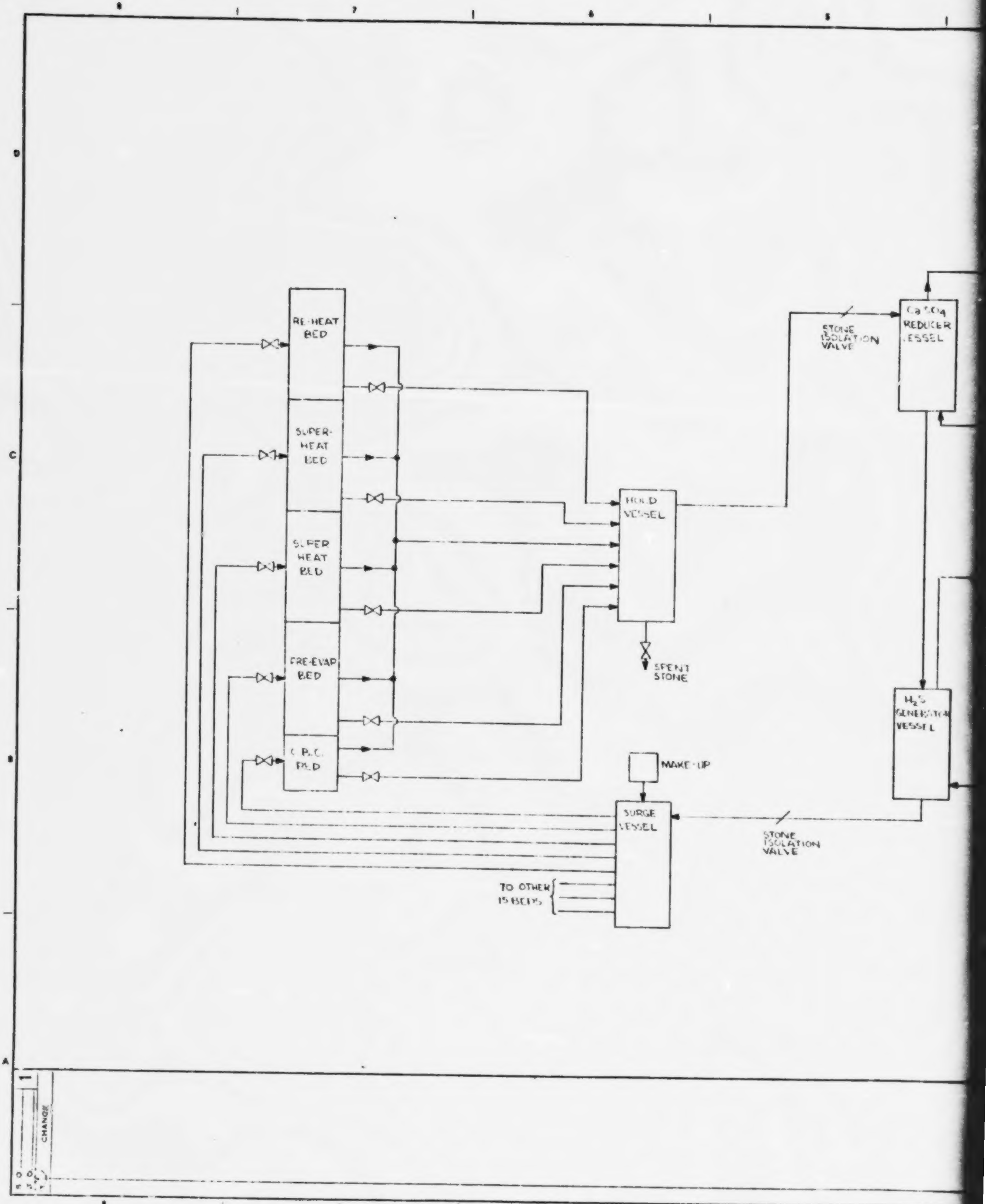
RE-HEAT BED
SUPER-HEAT BED
SUPER HEAT BED
PRE-EVAP BED
C.P.C. BED
HOLD VESSEL
SPENT STONE
MAKE-UP
SURGE VESSEL
TO OTHER 15 BEDS
STONE ISOLATION VALVE
CaSO4 REDUCER VESSEL
STONE ISOLATION VALVE
H2S GENERATOR VESSEL
CHANGE
8
7
6
5
D
C
B
A

A

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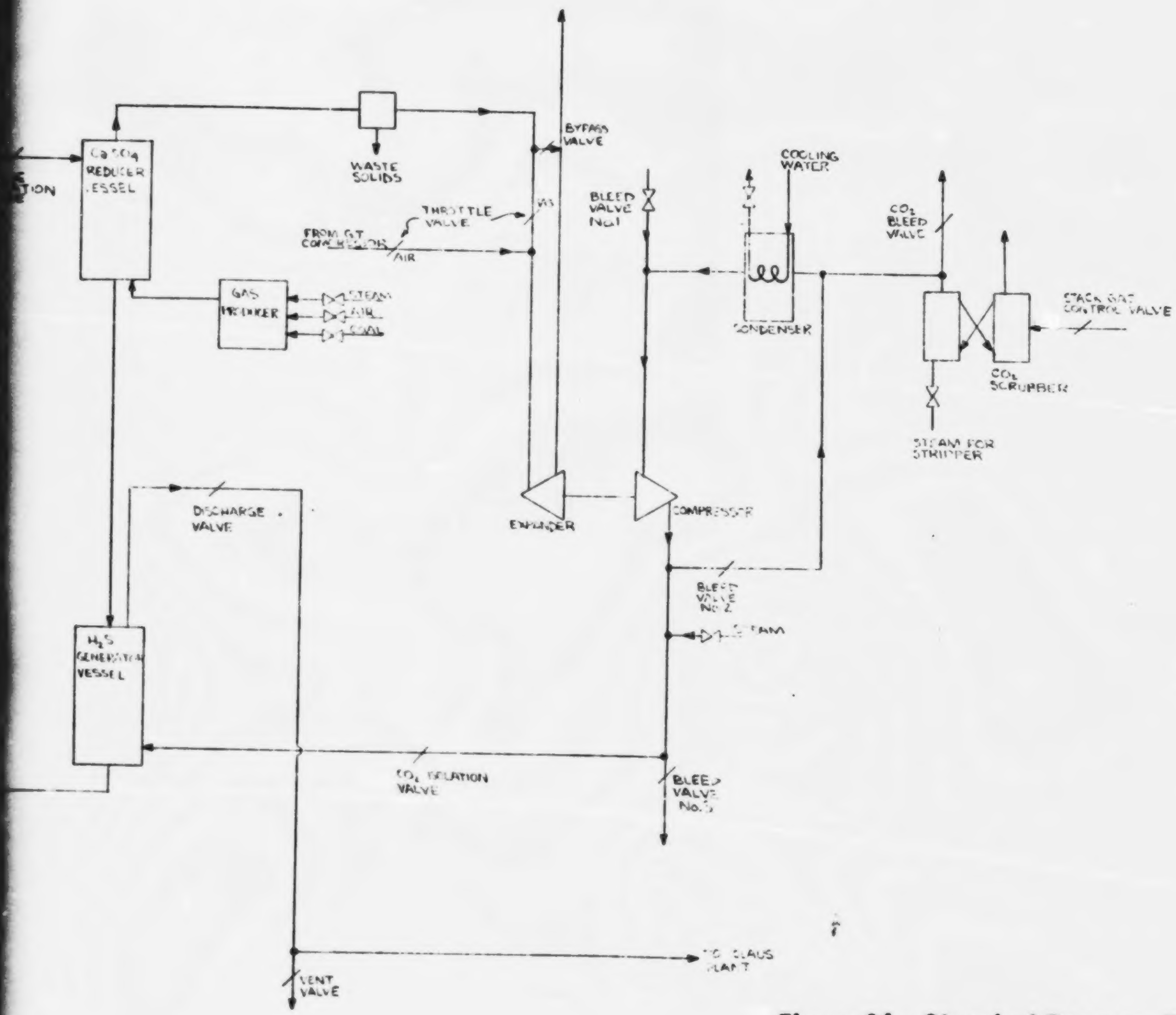
CaSO4 REDUCER VESSEL
WASTE SOLIDS
BYPASS VALVE
THROTTLE VALVE
FROM G.T. COMPRESSOR
AIR
GAS PRODUCER
STEAM
AIR
COAL
BLEED VALVE No.1
COOLING WATER
CONDENSER
CO2 BLEED VALVE
STACK GAS CONTROL VALVE
CO2 SCRUBBER
STEAM FOR STRIPPER
EXPANDER
COMPRESSOR
DISCHARGE VALVE
BLEED VALVE No.2
STEAM
H2S GENERATOR VESSEL
CO2 SEPARATION VALVE
BLEED VALVE No.3
TO CLAUS PLANT
VENT VALVE

Figure 94: Chemical Recovery Plant

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COMPUTER AND INSTRUMENTATION DIVISION

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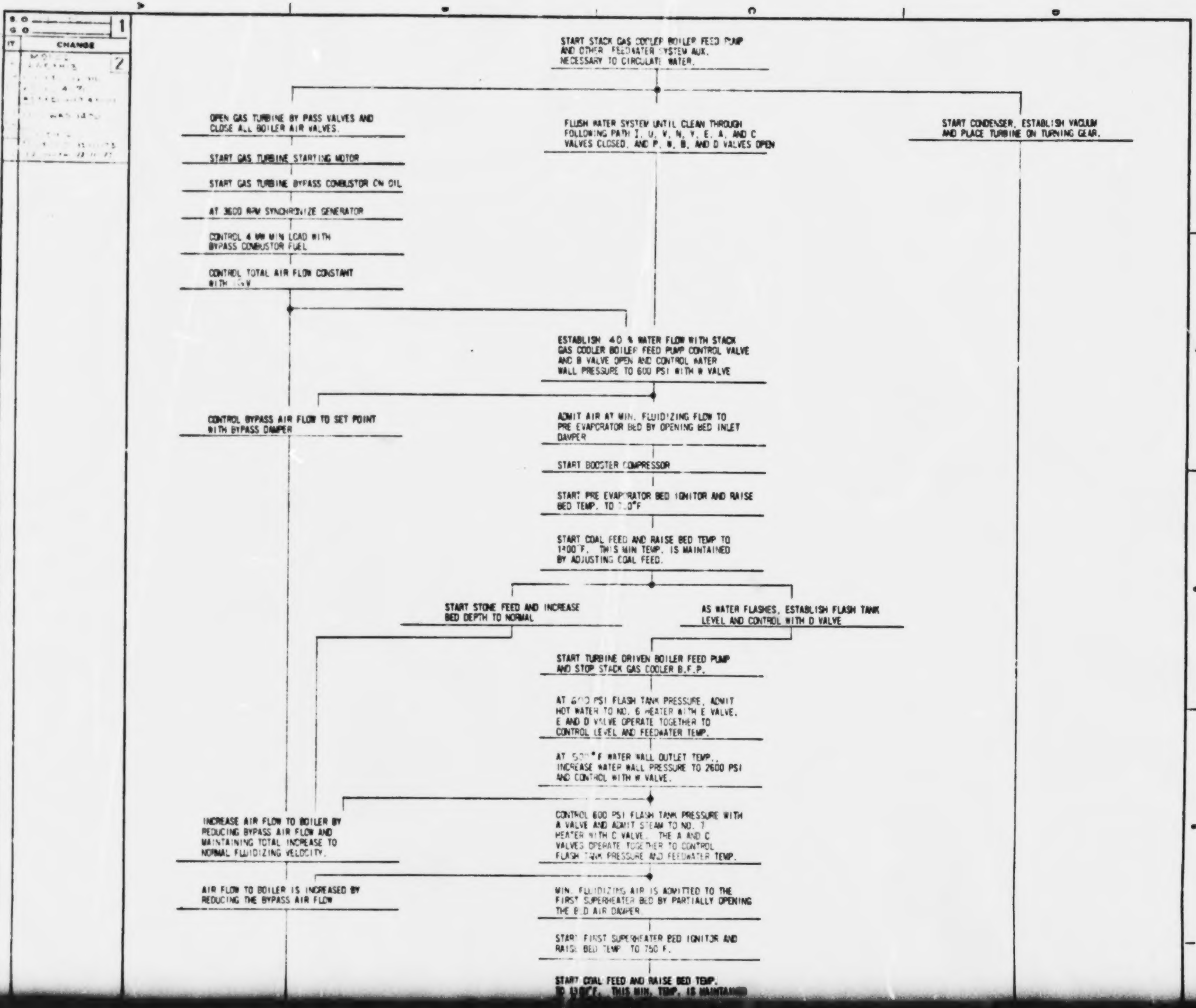
CHANGE
1
2
A
B
C
D
START STACK GAS COOLER BOILER FEED PUMP AND OTHER FEEDWATER SYSTEM AUX. NECESSARY TO CIRCULATE WATER.
OPEN GAS TURBINE BY PASS VALVES AND CLOSE ALL BOILER AIR VALVES.
FLUSH WATER SYSTEM UNTIL CLEAN THROUGH FOLLOWING PATH I, U, V, N, Y, E, A, AND C VALVES CLOSED, AND P, W, B, AND D VALVES OPEN
START CONDENSER, ESTABLISH VACUUM AND PLACE TURBINE ON TURNING GEAR.
START GAS TURBINE STARTING MOTOR
START GAS TURBINE BYPASS COMBUSTOR ON OIL
AT 3600 RPM SYNCHRONIZE GENERATOR
CONTROL 4 MW MIN LOAD WITH BYPASS COMBUSTOR FUEL
CONTROL TOTAL AIR FLOW CONSTANT WITH IGV
ESTABLISH 40 % WATER FLOW WITH STACK GAS COOLER BOILER FEED PUMP CONTROL VALVE AND B VALVE OPEN AND CONTROL WATER WALL PRESSURE TO 600 PSI WITH W VALVE
CONTROL BYPASS AIR FLOW TO SET POINT WITH BYPASS DAMPER
ADMIT AIR AT MIN. FLUIDIZING FLOW TO PRE EVAPORATOR BED BY OPENING BED INLET DAMPER
START BOOSTER COMPRESSOR
START PRE EVAPORATOR BED IGNITOR AND RAISE BED TEMP. TO 700°F
START COAL FEED AND RAISE BED TEMP TO 1400°F. THIS MIN TEMP. IS MAINTAINED BY ADJUSTING COAL FEED.
START STONE FEED AND INCREASE BED DEPTH TO NORMAL
AS WATER FLASHES, ESTABLISH FLASH TANK LEVEL AND CONTROL WITH D VALVE
START TURBINE DRIVEN BOILER FEED PUMP AND STOP STACK GAS COOLER B.F.P.
AT 600 PSI FLASH TANK PRESSURE, ADMIT HOT WATER TO NO. 6 HEATER WITH E VALVE. E AND D VALVE OPERATE TOGETHER TO CONTROL LEVEL AND FEEDWATER TEMP.
AT 500°F WATER WALL OUTLET TEMP., INCREASE WATER WALL PRESSURE TO 2600 PSI AND CONTROL WITH W VALVE.
INCREASE AIR FLOW TO BOILER BY REDUCING BYPASS AIR FLOW AND MAINTAINING TOTAL INCREASE TO NORMAL FLUIDIZING VELOCITY.
CONTROL 600 PSI FLASH TANK PRESSURE WITH A VALVE AND ADMIT STEAM TO NO. 7 HEATER WITH C VALVE. THE A AND C VALVES OPERATE TOGETHER TO CONTROL FLASH TANK PRESSURE AND FEEDWATER TEMP.
AIR FLOW TO BOILER IS INCREASED BY REDUCING THE BYPASS AIR FLOW
MIN. FLUIDIZING AIR IS ADMITTED TO THE FIRST SUPERHEATER BED BY PARTIALLY OPENING THE BED AIR DAMPER.
START FIRST SUPERHEATER BED IGNITOR AND RAISE BED TEMP. TO 750 F.
START COAL FEED AND RAISE BED TEMP. TO 1300°F. THIS MIN. TEMP. IS MAINTAI

A

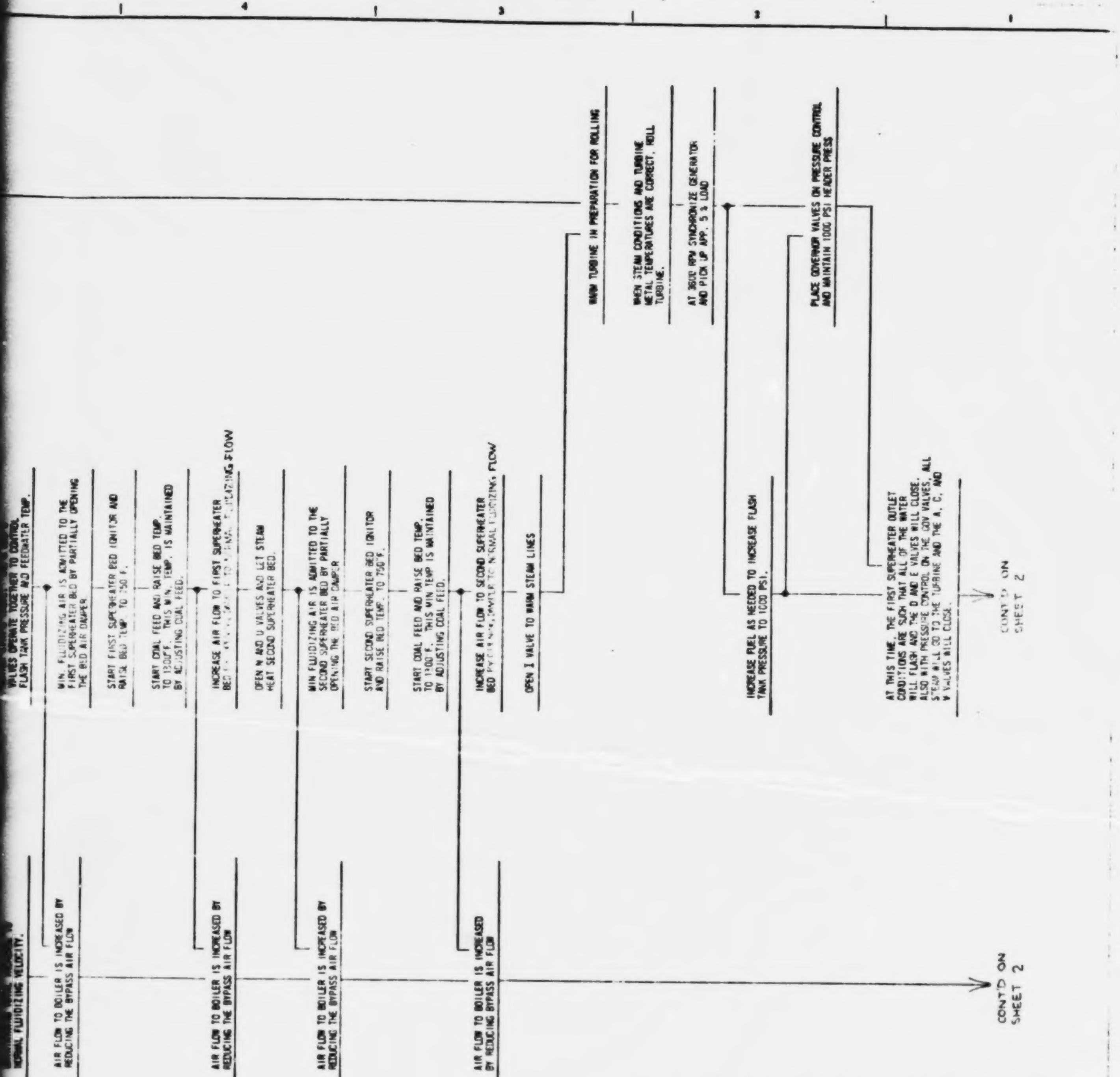
NORMAL FLUIDIZING VELOCITY.
VALVES OPERATE TOGETHER TO CONTROL FLASH TANK PRESSURE AND FEEDWATER TEMP.
AIR FLOW TO BOILER IS INCREASED BY REDUCING THE BYPASS AIR FLOW
MIN. FLUIDIZING AIR IS ADMITTED TO THE FIRST SUPERHEATER BED BY PARTIALLY OPENING THE BED AIR DAMPER
START FIRST SUPERHEATER BED IGNITOR AND RAISE BED TEMP. TO 750 F.
START COAL FEED AND RAISE BED TEMP. TO 1200 F. THIS MIN. TEMP. IS MAINTAINED BY ADJUSTING COAL FEED.
AIR FLOW TO BOILER IS INCREASED BY REDUCING THE BYPASS AIR FLOW
INCREASE AIR FLOW TO FIRST SUPERHEATER BED TO NORMAL FLUIDIZING FLOW
OPEN N AND U VALVES AND LET STEAM HEAT SECOND SUPERHEATER BED.
AIR FLOW TO BOILER IS INCREASED BY REDUCING THE BYPASS AIR FLOW
MIN FLUIDIZING AIR IS ADMITTED TO THE SECOND SUPERHEATER BED BY PARTIALLY OPENING THE BED AIR DAMPER
START SECOND SUPERHEATER BED IGNITOR AND RAISE BED TEMP. TO 750°F.
START COAL FEED AND RAISE BED TEMP. TO 1200 F. THIS MIN TEMP IS MAINTAINED BY ADJUSTING COAL FEED.
AIR FLOW TO BOILER IS INCREASED BY REDUCING BYPASS AIR FLOW
INCREASE AIR FLOW TO SECOND SUPERHEATER BED PYT... TO NORMAL FLUIDIZING FLOW
OPEN I VALVE TO WARM STEAM LINES
WARM TURBINE IN PREPARATION FOR ROLLING
WHEN STEAM CONDITIONS AND TURBINE METAL TEMPERATURES ARE CORRECT, ROLL TURBINE.
AT 3600 RPM SYNCHRONIZE GENERATOR AND PICK UP APP. 5 % LOAD
INCREASE FUEL AS NEEDED TO INCREASE FLASH TANK PRESSURE TO 1000 PSI.
PLACE GOVERNOR VALVES ON PRESSURE CONTROL AND MAINTAIN 1000 PSI HEADER PRESS
AT THIS TIME, THE FIRST SUPERHEATER OUTLET CONDITIONS ARE SUCH THAT ALL OF THE WATER WILL FLASH AND THE D AND E VALVES WILL CLOSE. ALSO WITH PRESSURE CONTROL ON THE GOV VALVES, ALL STEAM WILL GO TO THE TURBINE AND THE A, C, AND V VALVES WILL CLOSE.
CONT'D ON SHEET 2
CONT'D ON SHEET 2
1
4
1
3
1
2
1
1

Figure 95A: First Boiler Module Start-Up

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TITLE FIRST BOILER MODULE START UP

DIMENSIONS IN INCHES SCALE

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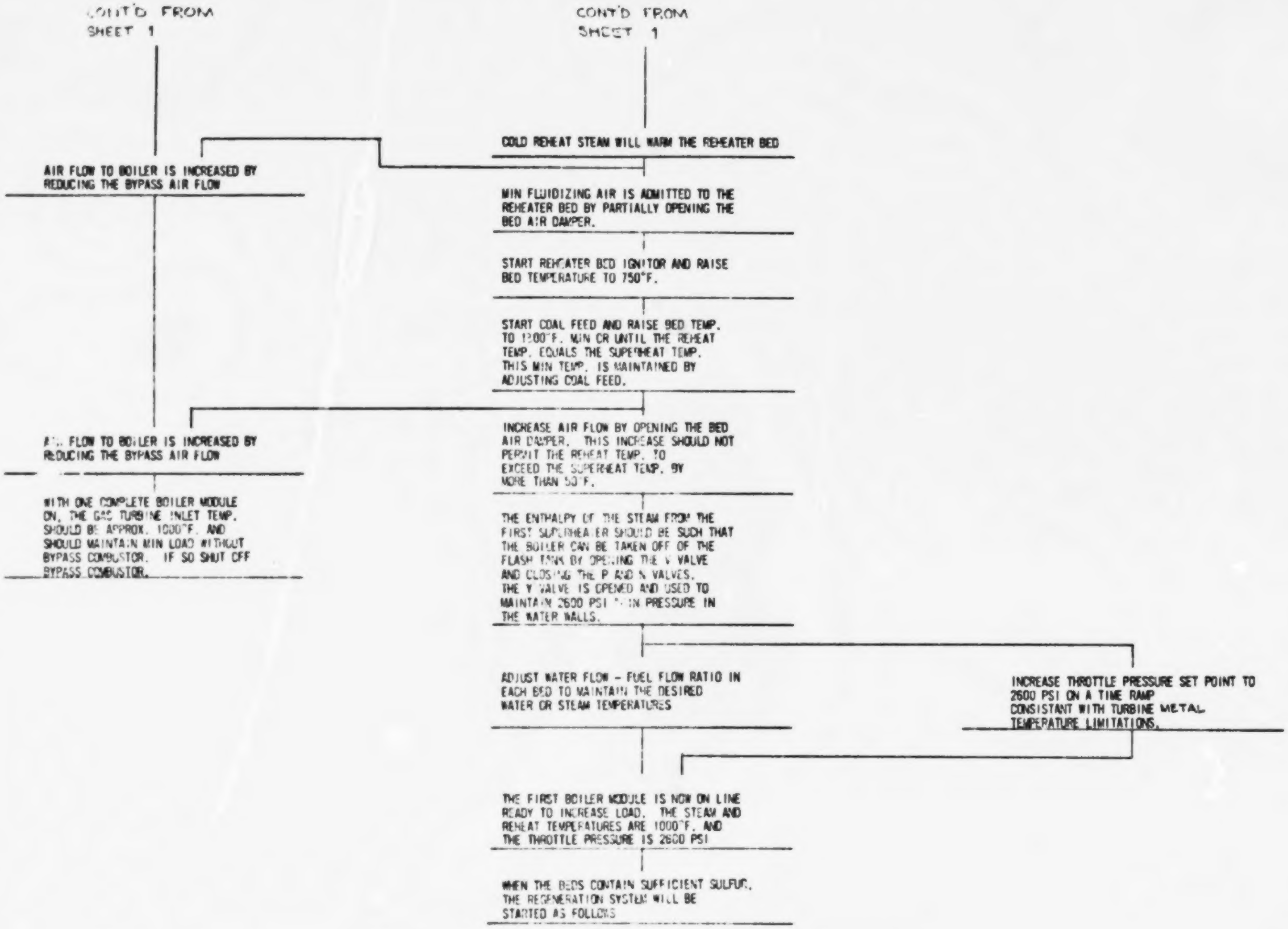

AIR FLOW TO BOILER IS INCREASED BY REDUCING THE BYPASS AIR FLOW

AIR FLOW TO BOILER IS INCREASED BY REDUCING THE BYPASS AIR FLOW

WITH ONE COMPLETE BOILER MODULE ON, THE GAS TURBINE INLET TEMP. SHOULD BE APPROX. 1000°F. AND SHOULD MAINTAIN MIN LOAD WITHOUT BYPASS COMBUSTOR. IF SO SHUT OFF BYPASS COMBUSTOR.

COLD REHEAT STEAM WILL WARM THE REHEATER BED

MIN FLUIDIZING AIR IS ADMITTED TO THE REHEATER BED BY PARTIALLY OPENING THE BED AIR DAMPER.

START REHEATER BED IGNITOR AND RAISE BED TEMPERATURE TO 750°F.

START COAL FEED AND RAISE BED TEMP. TO 1200°F. MIN OR UNTIL THE REHEAT TEMP. EQUALS THE SUPERHEAT TEMP. THIS MIN TEMP. IS MAINTAINED BY ADJUSTING COAL FEED.

INCREASE AIR FLOW BY OPENING THE BED AIR DAMPER. THIS INCREASE SHOULD NOT PERMIT THE REHEAT TEMP. TO EXCEED THE SUPERHEAT TEMP. BY MORE THAN 50°F.

THE ENTHALPY OF THE STEAM FROM THE FIRST SUPERHEATER SHOULD BE SUCH THAT THE BOILER CAN BE TAKEN OFF OF THE FLASH TANK BY OPENING THE V VALVE AND CLOSING THE P AND N VALVES. THE V VALVE IS OPENED AND USED TO MAINTAIN 2600 PSI MIN PRESSURE IN THE WATER WALLS.

ADJUST WATER FLOW - FUEL FLOW RATIO IN EACH BED TO MAINTAIN THE DESIRED WATER OR STEAM TEMPERATURES

INCREASE THROTTLE PRESSURE SET POINT TO 2600 PSI ON A TIME RAMP CONSISTANT WITH TURBINE METAL TEMPERATURE LIMITATIONS.

THE FIRST BOILER MODULE IS NOW ON LINE READY TO INCREASE LOAD. THE STEAM AND REHEAT TEMPERATURES ARE 1000°F. AND THE THROTTLE PRESSURE IS 2600 PSI

WHEN THE BEDS CONTAIN SUFFICIENT SULFUR, THE REGENERATION SYSTEM WILL BE STARTED AS FOLLOWS

Figure 95B: First Boiler Module Start-Up

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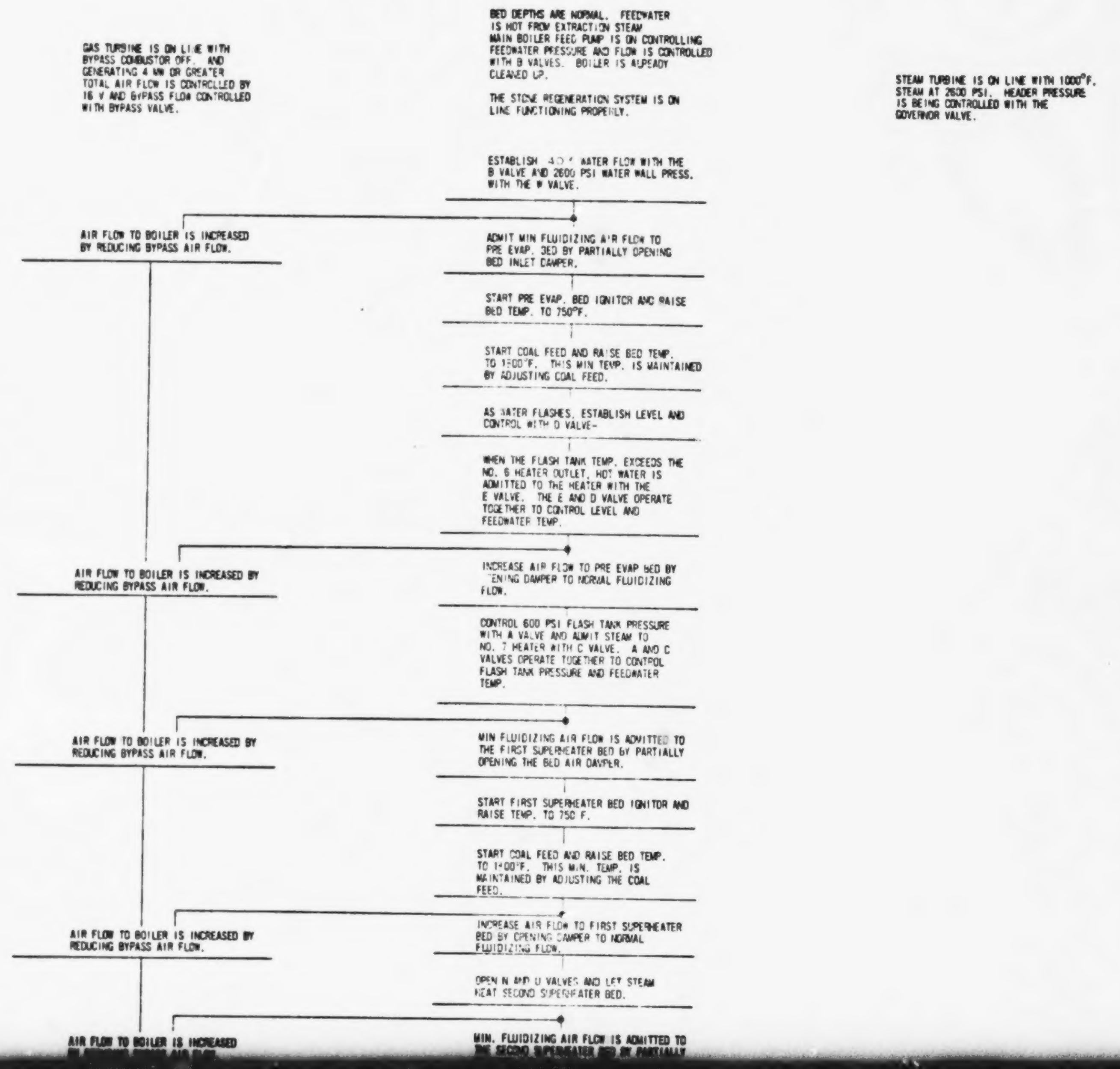

GAS TURBINE IS ON LINE WITH BYPASS COMBUSTOR OFF. AND GENERATING 4 MW OR GREATER TOTAL AIR FLOW IS CONTROLLED BY 16 V AND BYPASS FLOW CONTROLLED WITH BYPASS VALVE.

BED DEPTHS ARE NORMAL. FEEDWATER IS HOT FROM EXTRACTION STEAM MAIN BOILER FEED PUMP IS ON CONTROLLING FEEDWATER PRESSURE AND FLOW IS CONTROLLED WITH B VALVES. BOILER IS ALREADY CLEANED UP.

THE STONE REGENERATION SYSTEM IS ON LINE FUNCTIONING PROPERLY.

STEAM TURBINE IS ON LINE WITH 1000°F. STEAM AT 2600 PSI. HEADER PRESSURE IS BEING CONTROLLED WITH THE GOVERNOR VALVE.

ESTABLISH 40 % WATER FLOW WITH THE B VALVE AND 2600 PSI WATER WALL PRESS. WITH THE W VALVE.

AIR FLOW TO BOILER IS INCREASED BY REDUCING BYPASS AIR FLOW.

ADMIT MIN FLUIDIZING AIR FLOW TO PRE EVAP. BED BY PARTIALLY OPENING BED INLET DAMPER.

START PRE EVAP. BED IGNITOR AND RAISE BED TEMP. TO 750°F.

START COAL FEED AND RAISE BED TEMP. TO 1500°F. THIS MIN TEMP. IS MAINTAINED BY ADJUSTING COAL FEED.

AS WATER FLASHES, ESTABLISH LEVEL AND CONTROL WITH D VALVE.

WHEN THE FLASH TANK TEMP. EXCEEDS THE NO. 6 HEATER OUTLET, HOT WATER IS ADMITTED TO THE HEATER WITH THE E VALVE. THE E AND D VALVE OPERATE TOGETHER TO CONTROL LEVEL AND FEEDWATER TEMP.

AIR FLOW TO BOILER IS INCREASED BY REDUCING BYPASS AIR FLOW.

INCREASE AIR FLOW TO PRE EVAP BED BY OPENING DAMPER TO NORMAL FLUIDIZING FLOW.

CONTROL 600 PSI FLASH TANK PRESSURE WITH A VALVE AND ADMIT STEAM TO NO. 7 HEATER WITH C VALVE. A AND C VALVES OPERATE TOGETHER TO CONTROL FLASH TANK PRESSURE AND FEEDWATER TEMP.

AIR FLOW TO BOILER IS INCREASED BY REDUCING BYPASS AIR FLOW.

MIN FLUIDIZING AIR FLOW IS ADMITTED TO THE FIRST SUPERHEATER BED BY PARTIALLY OPENING THE BED AIR DAMPER.

START FIRST SUPERHEATER BED IGNITOR AND RAISE TEMP. TO 750 F.

START COAL FEED AND RAISE BED TEMP. TO 1400°F. THIS MIN. TEMP. IS MAINTAINED BY ADJUSTING THE COAL FEED.

AIR FLOW TO BOILER IS INCREASED BY REDUCING BYPASS AIR FLOW.

INCREASE AIR FLOW TO FIRST SUPERHEATER BED BY OPENING DAMPER TO NORMAL FLUIDIZING FLOW.

OPEN N AND U VALVES AND LET STEAM HEAT SECOND SUPERHEATER BED.

AIR FLOW TO BOILER IS INCREASED

MIN. FLUIDIZING AIR FLOW IS ADMITTED TO THE SECOND SUPERHEATER BED BY PARTIALLY

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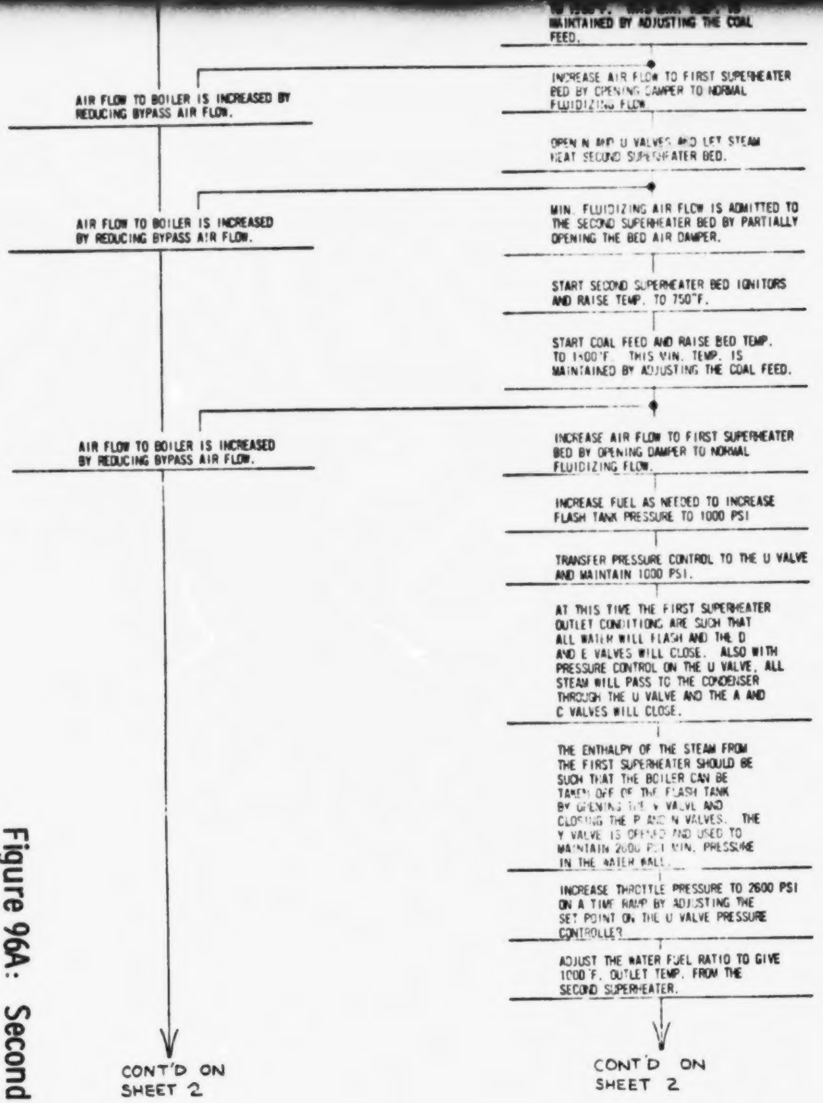

MAINTAINED BY ADJUSTING THE COAL FEED.

INCREASE AIR FLOW TO FIRST SUPERHEATER BED BY OPENING DAMPER TO NORMAL FLUIDIZING FLOW.

OPEN N AND U VALVES AND LET STEAM HEAT SECOND SUPERHEATER BED.

MIN. FLUIDIZING AIR FLOW IS ADMITTED TO THE SECOND SUPERHEATER BED BY PARTIALLY OPENING THE BED AIR DAMPER.

START SECOND SUPERHEATER BED IGNITORS AND RAISE TEMP. TO 750°F.

START COAL FEED AND RAISE BED TEMP. TO 1400°F. THIS MIN. TEMP. IS MAINTAINED BY ADJUSTING THE COAL FEED.

INCREASE AIR FLOW TO FIRST SUPERHEATER BED BY OPENING DAMPER TO NORMAL FLUIDIZING FLOW.

INCREASE FUEL AS NEEDED TO INCREASE FLASH TANK PRESSURE TO 1000 PSI

TRANSFER PRESSURE CONTROL TO THE U VALVE AND MAINTAIN 1000 PSI.

AT THIS TIME THE FIRST SUPERHEATER OUTLET CONDITIONS ARE SUCH THAT ALL WATER WILL FLASH AND THE D AND E VALVES WILL CLOSE. ALSO WITH PRESSURE CONTROL ON THE U VALVE, ALL STEAM WILL PASS TO THE CONDENSER THROUGH THE U VALVE AND THE A AND C VALVES WILL CLOSE.

THE ENTHALPY OF THE STEAM FROM THE FIRST SUPERHEATER SHOULD BE SUCH THAT THE BOILER CAN BE TAKEN OFF OF THE FLASH TANK BY OPENING THE Y VALVE AND CLOSING THE P AND N VALVES. THE Y VALVE IS OPENED AND USED TO MAINTAIN 2000 PSI MIN. PRESSURE IN THE WATER WALL.

INCREASE THROTTLE PRESSURE TO 2600 PSI ON A TIME RAMP BY ADJUSTING THE SET POINT ON THE U VALVE PRESSURE CONTROLLER.

ADJUST THE WATER FUEL RATIO TO GIVE 1000°F. OUTLET TEMP. FROM THE SECOND SUPERHEATER.

AIR FLOW TO BOILER IS INCREASED BY REDUCING BYPASS AIR FLOW.

AIR FLOW TO BOILER IS INCREASED BY REDUCING BYPASS AIR FLOW.

AIR FLOW TO BOILER IS INCREASED BY REDUCING BYPASS AIR FLOW.

Figure 96A: Second Boiler Module Start-Up

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TITLE SECOND BOILER MODULE START UP
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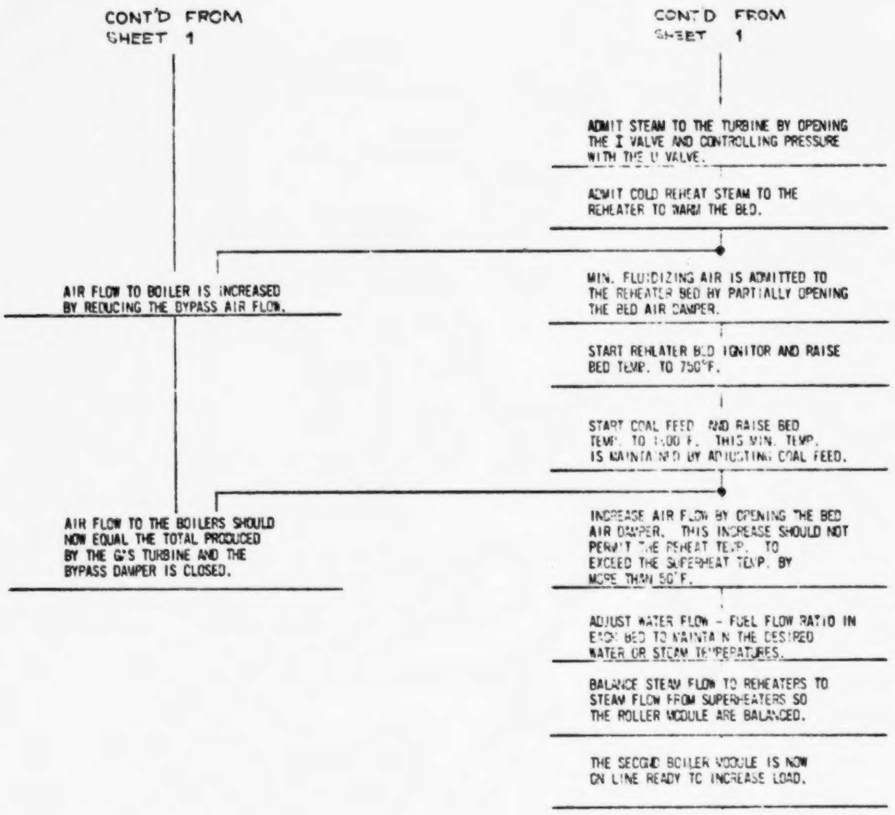

AIR FLOW TO BOILER IS INCREASED BY REDUCING THE BYPASS AIR FLOW.

AIR FLOW TO THE BOILERS SHOULD NOW EQUAL THE TOTAL PRODUCED BY THE G'S TURBINE AND THE BYPASS DAMPER IS CLOSED.

ADMIT STEAM TO THE TURBINE BY OPENING THE I VALVE AND CONTROLLING PRESSURE WITH THE U VALVE.

ADMIT COLD REHEAT STEAM TO THE REHEATER TO WARM THE BED.

MIN. FLUIDIZING AIR IS ADMITTED TO THE REHEATER BED BY PARTIALLY OPENING THE BED AIR DAMPER.

START REHEATER BED IGNITOR AND RAISE BED TEMP. TO 750°F.

START COAL FEED AND RAISE BED TEMP. TO 1100 F. THIS MIN. TEMP. IS MAINTAINED BY ADJUSTING COAL FEED.

INCREASE AIR FLOW BY OPENING THE BED AIR DAMPER. THIS INCREASE SHOULD NOT PERMIT THE REHEAT TEMP. TO EXCEED THE SUPERHEAT TEMP. BY MORE THAN 50°F.

ADJUST WATER FLOW - FUEL FLOW RATIO IN EACH BED TO MAINTAIN THE DESIRED WATER OR STEAM TEMPERATURES.

BALANCE STEAM FLOW TO REHEATERS TO STEAM FLOW FROM SUPERHEATERS SO THE ROLLER MODULE ARE BALANCED.

THE SECOND BOILER MODULE IS NOW ON LINE READY TO INCREASE LOAD.

Figure 96B: Second Boiler Module Start-Up

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5.2.2 First-Module Start-Up

The start-up procedure for the first module assumes that the plant is cold and completely off-line. Only the starting of the gas turbine, boiler module, steam turbine, and sulfur recovery system are covered in this description. Starting the condenser, getting the steam turbine on turning gear, filling the system with water, flushing the water system, and other conventional start-up procedures are not covered in this description as they are the same as those for a conventional subcritical once-through steam plant.

The gas turbine will be started in a manner similar to that for starting a conventional gas turbine, except that this unit utilizes a separate combustion chamber. The combustion system will be considerably smaller than normal. This will reduce the initial temperature overshoot and make the start-up smooth, with little danger of surge or high turbine inlet temperatures. The gas-turbine start-up will proceed in the normal manner with all of the compressor airflow passing through the by-pass combustor until the turbine has reached synchronous speed and has been synchronized. The minimum load of approximately 4 MW will be controlled during the boiler module start-up by adjusting the gas-turbine combustor fuel flow. The stack gas cooler boiler feed pump will be started and a flow of 40% of boiler module capacity established. For the first module, the flow will be controlled with the stack gas cooler boiler feed pump valve, and the B valve on the module will be open. Air will be admitted to the preevaporator module by opening the module inlet damper, starting the booster air compressor, and throttling the by-pass airflow. The flow rate will be controlled by positioning the by-pass damper. The bed will be filled with fresh stone at a reduced depth, and the airflow will be at a minimum value so that the heat required by the ignitors will be held to a minimum. The preevaporator igniters will be started and the bed temperature increased to approximately 700°F, the temperature required to ignite the coal.

Coal feed will be started and bed temperature increased to 1300°F. The bed depth and airflow are then increased to the normal level.

When the water wall outlet temperature has increased sufficiently, air will be admitted to the first superheater bed and the igniters started. The first superheater bed will also be filled with clean stone at a reduced depth. Coal feed will be started at ignition temperature and the bed temperature increased to 1300°F. The bed depth and airflow are then increased to the normal levels.

During this time, the gas-turbine combustor fuel will be modulated to maintain a minimum load of 4 MW on the gas-turbine generator. Airflow to the boiler modules will be controlled by positioning the by-pass valve; and total airflow will be controlled by throttling the inlet guide vanes on the gas turbine as required.

The start-up progresses as described previously in "Evaluation of Fluidized Bed Combustion,"Vol.III, Westinghouse Research Laboratories report to Office of Air Programs, EPA, Nov. 1971.

Several operating points are not mentioned in the start-up procedure of that report. These include the fact that when the first-stage superheater outlet temperature is sufficiently high, a turbine-driven boiler feed pump will be started and the feedwater passed through the feedwater heaters. This permits heat recovery in the no. 6 and 7 heaters. The stack gas cooler boiler feed pump will be shut down at this time.

The reheater will be fired prior to admitting steam to the turbine, with the bed temperature limited to 880°F. When steam is admitted to the reheater, the reheat temperature set point will be set equal to the superheat steam value. By this means, the reheat and superheat steam temperature will be the same and the temperature shock to the turbine will be minimized. The boiler module outlet temperatures are sufficient to permit shutting off the gas-turbine combustor when the reheater is fired.

A desuperheater will be interlocked with the U valve so steam can be sent to the condenser.

5.2.3 Second-Module Start-Up

The second boiler module start-up is similar to that for the first boiler module except that the first boiler exit temperature will be sufficient to permit the by-pass combustors to be shut down and the feedwater will be hot. These differences will cause the following changes to the procedure described for the first unit:

- Start-up of a gas turbine will not be required.
- The depth of the preevaporator and first superheater beds will not need to be reduced for ignition.
- The stack gas cooler boiler feed pump will not be used, as water flow will be through the feedwater heaters; feedwater flow will be controlled by positioning the B valve.

Since the feedwater temperature will be considerably higher than the 80°F of the first module, the heat-recovery portion of the start-up cycle will be reduced. The E and C valves will be interlocked so they will not admit water or steam from the flash tank until the flash tank temperature exceeds the output temperature from the no. 6 heater for the E valve and the no. 7 heater for the C valve.

The I valve cannot be opened until the superheater pressure reaches 2600 psi and the superheater outlet temperature is 1000°F. Therefore, the system must be pressurized by use of the U valve. This process will proceed faster than that for the first unit because no steam temperature changes to the turbine will be involved.

The reheater light-off can be delayed because the steam can be diverted to the first boiler module reheater. Also, this module can be preheated with steam from the turbine prior to light-off. These features make the light-off of this reheater module less critical than that of the first module reheater.

The turbine load will be increased to equal the water flow through the B valve by opening the I valve. The superheater outlet pressure will be maintained by modulating the U valve. After the I valve is open, the pressure set point for the U valve will be increased to serve as a relief valve.

When the I valve is 100 percent open, the load will be increased further by increasing the flow through the B valve. The water flow must now be controlled by modulating the B valves and holding the feedwater heater pressure constant by varying the boiler feed pump speed.

The steam flow through the reheaters must be balanced or maintained proportional to the flow through the B valves by positioning the reheater inlet valves.

5.2.4 Third-Module Start-up

The start of the third module will be the same as for starting the gas turbine and airflow system for the first boiler module and for starting the second boiler module, the difference being the balance and comparisons of three boiler modules instead of two.

5.2.5 Regeneration System Start-up

The plant will be started up by operating the stone feed once-through rather than regenerating. This will be accomplished by starting with clean stone in the beds, with the surge vessel full of clean stone, and with the hold vessel empty. No stone will be added to or removed from the beds until SO_2 is sensed at the boiler outlet. Clean stone will be added to the bed producing the SO_2 from the surge vessel. The spent stone will be removed and placed in the hold vessel. The boiler module can operate in this mode until the hold vessel is full. This provides sufficient time for the plant operation to stabilize and permits starting the chemical regeneration system without any adverse effects on the remainder of the plant.

When starting the specific regeneration system is desired, as shown in Figure 94, the $CaSO_4$ reducer vessel and the H_2S generator vessel will be filled with stone from the hold vessel and the two stone isolation valves closed. The gas producer will be started at atmospheric pressure and the gas passed through the $CaSO_4$ reducer vessel and discharged through the by-pass valve to the stack. The gas throttle valve to the expander will be closed. The expander will be started with air from the gas turbine and the air throttle valve used for control. Air to the compressor will enter through bleed valve no. 1, which modulates to control the required flow and is discharged through bleed valve no. 3, which modulates to control the desired discharge pressure. The CO_2 isolation valve and bleed valve no. 2 will be closed. The CO_2 scrubber will be started by admitting the desired flow of stack gas. This will be controlled by modulating the stack gas control valve. The stripper flow will be controlled to maintain the desired temperature in the CO_2 scrubber. As CO_2 flow increases, bleed valve no. 1 will be closed and bleed valve no. 2 opened to maintain the desired minimum flow to the compressor. Cooling water will be started to the condenser to maintain the desired CO_2 temperature. When bleed valve no. 1 is closed and all of the air has been purged from the system, the surge control will be transferred to bleed valve no. 2 and bleed valve no. 3 closed. The CO_2 flow will be started to the H_2S generator by opening the CO_2 isolation valve and controlling the system pressure to approximately 2 atmospheres with the discharge valve. The system pressures will be increased with the discharge valve and by-pass valves to maintain the H_2S generator vessel pressure approximately 2 atmospheres above that of the $CaSO_4$ reducer vessel. Pressures will continue to increase until the $CaSO_4$ reducer vessel pressure is 2 atmospheres below those of the boiler module. The H_2O to CO_2 ratio will be established and maintained by admitting steam to the CO_2 compressor discharge. When the proper pressures are attained and the system stabilized, the two stone isolation valves will be opened to start the stone flow. The expander can now be transferred to the $CaSO_4$ reducer vessel discharge gas by transferring from the air throttle valve to the gas throttle valve.

The control systems will be transferred to the on-line mode from the start-up mode.

5.2.6 Auxiliary Fuel Storage Requirements

The estimated auxiliary fuel requirements for bringing four boiler modules on-line from a cold start are:

Gas turbine	15,500 lb
Igniters in beds	20,000 lb
Total	35,500 lb

For a hot start where three additional boiler modules are brought on-line from a minimum load condition with one boiler module, the estimated auxiliary fuel requirements are:

Gas turbine	6,000 lb
Igniters in beds	12,000 lb
Total	18,000 lb

Assuming no. 2 fuel oil with a specific gravity of 0.85, 5,000 gal. of fuel oil would be consumed in a cold start and 2,500 gal. would be consumed in a hot start.

Assuming intermediate load operation of the plant with one hot start per day and one cold start per week, the fuel storage capacity required for once a week delivery and 100% reserves would be 40,000 gal.

5.3 ON-LINE LOADING

5.3.1 Introduction

The start-up of the individual boiler modules is discussed in Section 5.2, but the discussion does not include when or in what order the boiler modules should be started.

The plant will contain four boiler modules, with a turndown range of approximately 2:1. That is, each can operate from 50% maximum load to maximum load. By operating the plant with various combinations of the four boiler modules, the plant load can be varied from 12-1/2% to 100% of maximum capacity.

The expected time to load the plant will be consistent with that of a large conventional fossil fired once-through boiler power plant. The time required for flash tank operation and pressurizing is the same for either type once-through boiler. For a cold start, the steam turbine requires a soak time of two to three hours at approximately 2200 rpm. This will permit time for the chemical process to stabilize and the auxiliary equipment to be placed in operation. The steam turbine requires a soak time of two to three hours at approximately 2200 rpm. This will permit time for the chemical process to stabilize and auxiliary equipment to be placed in operation. The steam turbine requires approximately 90 minutes to increase the load to 100% rated load from the time the generator is synchronized to the line. Proper manipulation of the second gas turbine and the three remaining boiler modules will permit the fluidized bed boiler to increase steam production as required to meet the steam-turbine needs for the 90-minute loading.

The plant load can be varied by adjusting the boiler modules simultaneously or sequentially. Sequential operation of the modules will require starting and stopping equipment. Simultaneous operation of the modules limits the load variation range to approximately 2:1. The fastest load changes are accomplished by adjusting the equipment in service. Five percent per minute load changes can be achieved by adjusting fuel and water to the modules. Time is required to start the modules. Load increases beyond the operating range of the equipment in service, therefore, will be slower than 5% per minute. Load decreases can be accomplished rapidly by tripping portions of the plant that is operating.

Three basic approaches to load changes will be discussed.

- Sequential loading
- Predictive loading
- Rapid loading.

5.3.2 Sequential Loading

The sequential-by-module loading scheme will be used for normal small increases or decreases as received from the load dispatcher. Figure 97 shows the sequential loading scheme on a load versus time (no time scale), for the total plant at a uniform rate and for the portion delivered by each module. When an increase or decrease request is received but the eventual end point is not known, the load on the first boiler module will be increased as required to 80%. If an additional load is requested, the second boiler module will be started. The steam flow from the first boiler module will be reduced as required to place the second boiler module on-line and not by-pass steam to the condenser. Additional load increases will be accomplished by increasing the first boiler module and keeping the second boiler module at minimum until the first boiler module is fully loaded. The second boiler module is then loaded as required to 80% load. If an additional load is requested, the third boiler module is started and placed on-line in the same manner as the second. The procedure is then repeated for the fourth boiler module.

This procedure will give the most efficient (but not the fastest) loading of the plant. In general, the best plant heat rate is obtained by operating with the least number of modules required to produce the desired load and to operate with all of the modules but one at high load.

The start-up of boiler modules requires the use of the flash tank, so the boiler modules must be started one at a time. The shutdown does not require the flash tank, so more than one boiler module can be taken off the line at the same time. The start-up or shutdown of the gas turbine can be done simultaneously with the starting or stopping of the boiler modules, if reduction of the total time is desired.

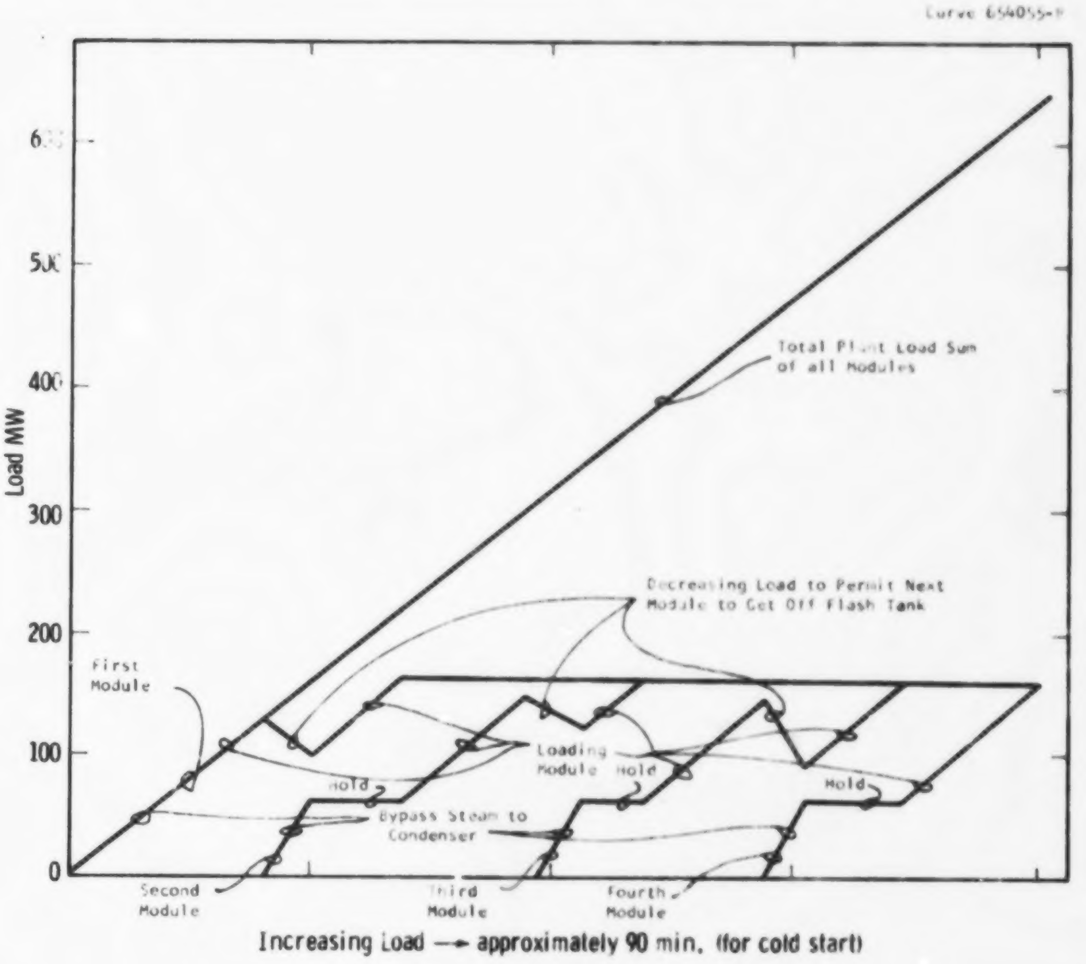


Figure 97: Sequential Loading Scheme

5.3.3 Predictive Loading

If the operator knows to what load he must have the plant at a given time, a program can determine the most desirable configuration of plant equipment and the best loading and starting sequence with which to achieve it. The program can operate for either an increasing or a decreasing load. The program will use but not be limited to the following items to determine the best loading and starting sequence:

- The best heat rate for each boiler module will be at high load.
- The reheaters can be started or stopped independently of the rest of the boiler modules.
- The boiler module temperature changes should be held to a minimum.
- The rate of change of the steam turbine is limited.
- The gas turbines can be started prior to being required and held at minimum load.
- The gas turbines must be run after module shutdown to regenerate the stone or lower the bed.
- The plant will contain only one flash tank and it can be used with only one boiler module at a time.
- The flash tank will not be needed for module shutdown, but the desuperheater will have limited capacity.
- Stone can be stored, made up, or operated once-through so the regeneration system will not be operated at low loads.
- Any configuration can be operated over its entire range, if necessary.

The predictive program will be used only when called for by the operator and when supplied with the proper information by the operator; otherwise, the sequential loading program will be used.

5.3.4 Rapid Loading

Plant load can be increased or decreased faster by operating the modules simultaneously or in parallel rather than sequentially. This method of operation can be used for the 2:1 operating range without by-passing steam to the condenser, or over a wider range by sending steam to the condenser.

Figure 98 shows the total operating range of each of the four configurations, including that portion accomplished by sending steam to the condenser. This figure shows that the operating range with four modules in service is from 265 MW to 620 MW; with three modules in service, 165 MW to 465 MW; with two modules in service, from 90 MW to 320 MW; and with one module in service, from 30 MW to 145 MW.

Load changes within these ranges can be accomplished at rates faster than 5% per minute. The plant will not be operating at its best heat rate under these conditions, so changes at such rates should be made only in an emergency and the load distribution adjusted slowly when the operation stabilizes to the more economical configuration.

5.3.5 Loading Events Diagram

The attached events diagram for a cold start of the fluidized bed boiler power plant shows the relation of pressures, temperatures, and flows at several points of the steam and water cycle and the air and gas cycle. The valves on the events diagram are not meant to be exact design data but to show graphically the simultaneous relation of the variables of the two plant cycles as the plant is started up.

Curve 654053-A

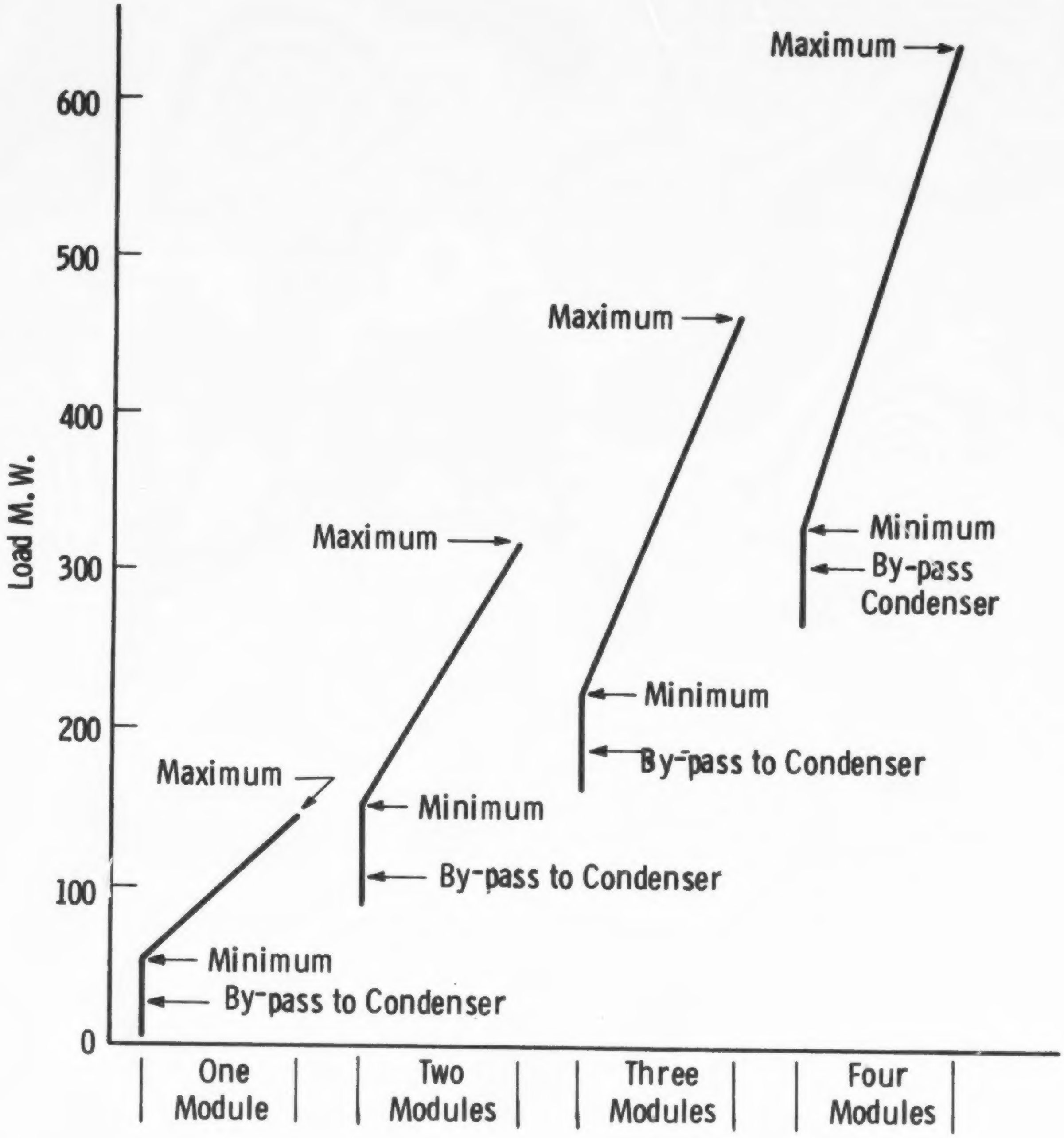


Figure 98: Module Operating Range

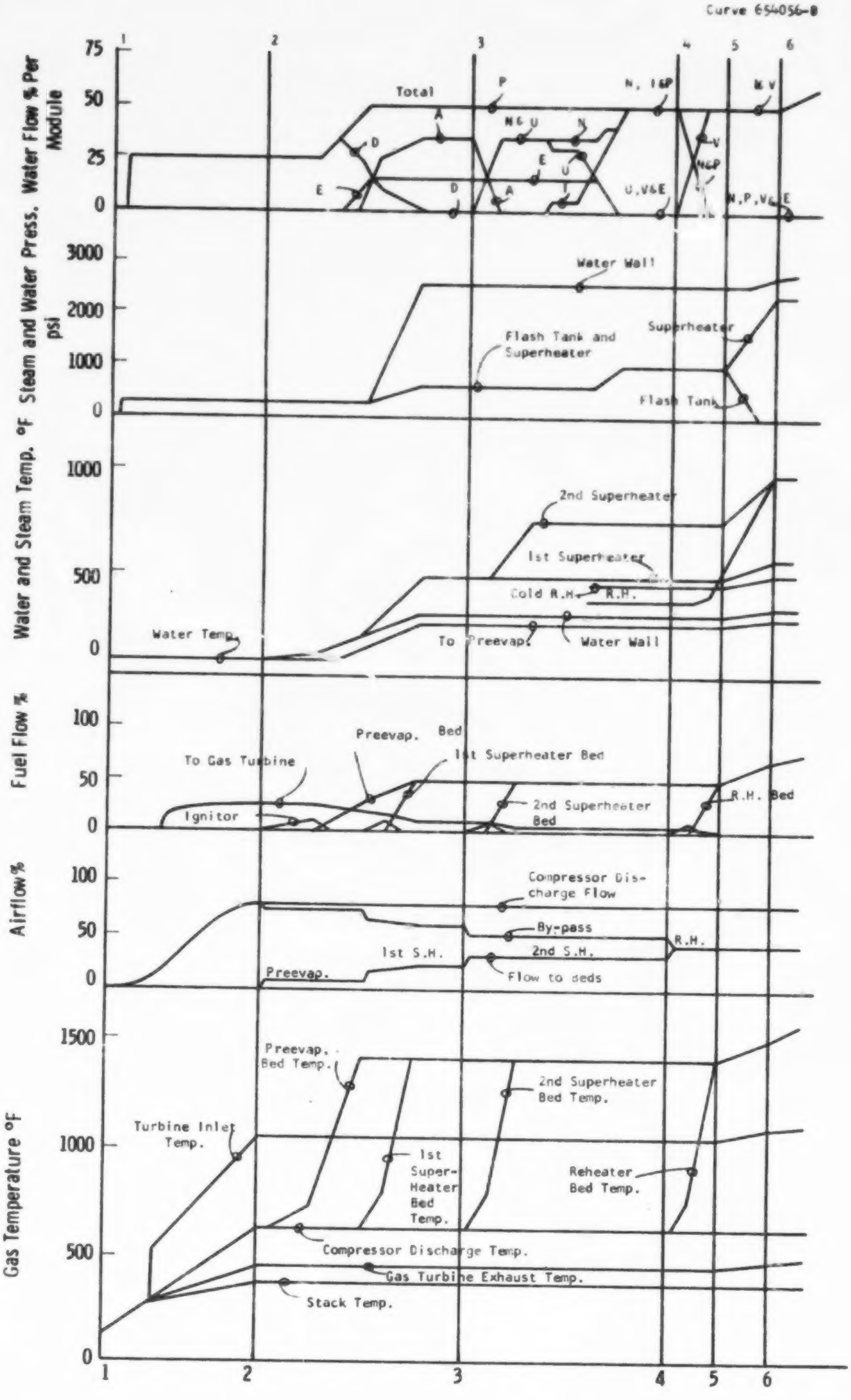


Figure 99—Event Diagram — Cold Start

The six events are as follows:

- The time the plant start-up begins
- The gas turbine has reached 3600 rpm and the first boiler bed ignitor is started.
- The second superheater is placed in service by closing the A valve, opening the N valve, and opening the V valve. This causes the steam to flow through the second superheater and then to the condenser, rather than from the flash tank to the condenser.
- The boiler is made once-through by opening the V valve and closing the N and P valves.
- The pressure ramp is started.
- The pressure ramp is completed and the boiler is at 2400 psi and ready to increase load.

5.3.6 Dynamic Analysis

An analysis was made to determine the ability of the fluidized bed boiler to change load. Two types of changes were made to the heat input: step changes of 5%, 10% and 15%; and ramp changes of 5% per minute, 10% per minute, and 15% per minute. These changes were imposed at the equivalent of 70% plant load. Figure 100 shows the bed temperature versus time for the two changes described above. Figure 101 shows the power output versus time for the two fuel input changes.

These curves show the expected effect on temperature and load caused by the stored energy in the beds. They also show that the unit is sensitive to step changes and to overfiring. The control system will utilize this information and overfeed fuel in predictable amounts to increase bed temperature as required to increase load, and the reverse to decrease load. If a nearly constant 5% per minute load increase is desired, the fuel rate will have to be increased by approximately 15% per minute for the first minute. This rate of change in fuel rate would

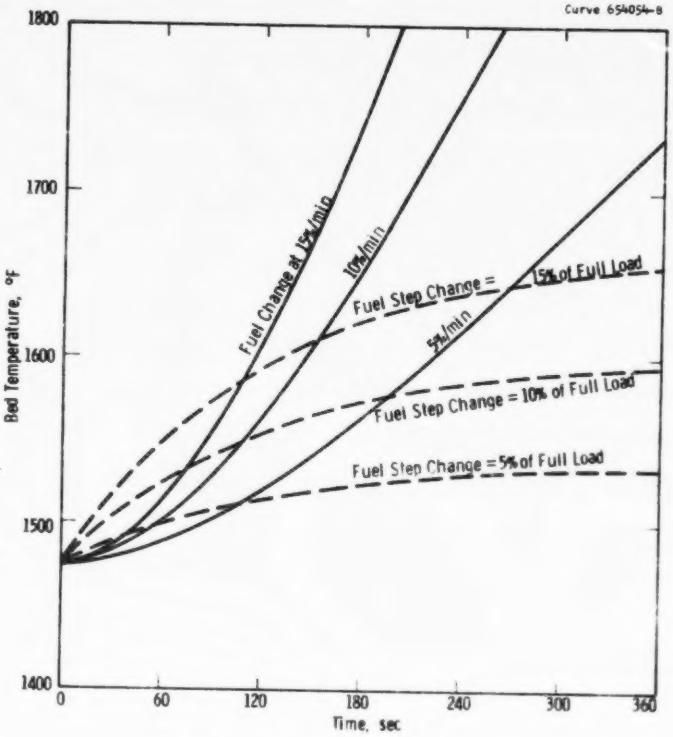


Figure 100: Bed Temperature Change as a Function of Fuel Feed Policy

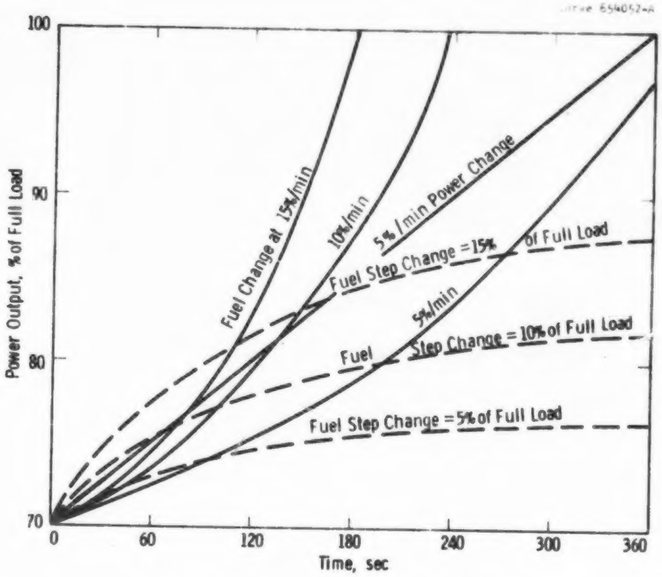


Figure 101: Load Response as a Function of Fuel Feed Policy

be decreased to 10% per minute by the end of the second minute, and to 7-1/2% per minute by the end of the fourth minute. Manipulating the fuel feed in this manner will permit the unit to change load at the desired 5% per minute.

5.4 UNIT SHUTDOWN

5.4.1 Introduction

Large load decreases will be accomplished by removing boiler modules and gas turbines from operation. Because of the complex configuration of the plant, several operating procedures are required:

- Shutting down the first boiler module associated with a gas turbine. The first or third modules to be shut down.
- Shutting down the second boiler module and one gas turbine
- Shutting down the fourth boiler module, one gas turbine, the steam turbine, and the chemical recovery plant.

Figures 93 and 94 are process schematics showing the control valves referred to in this section and their functional location in the plant.

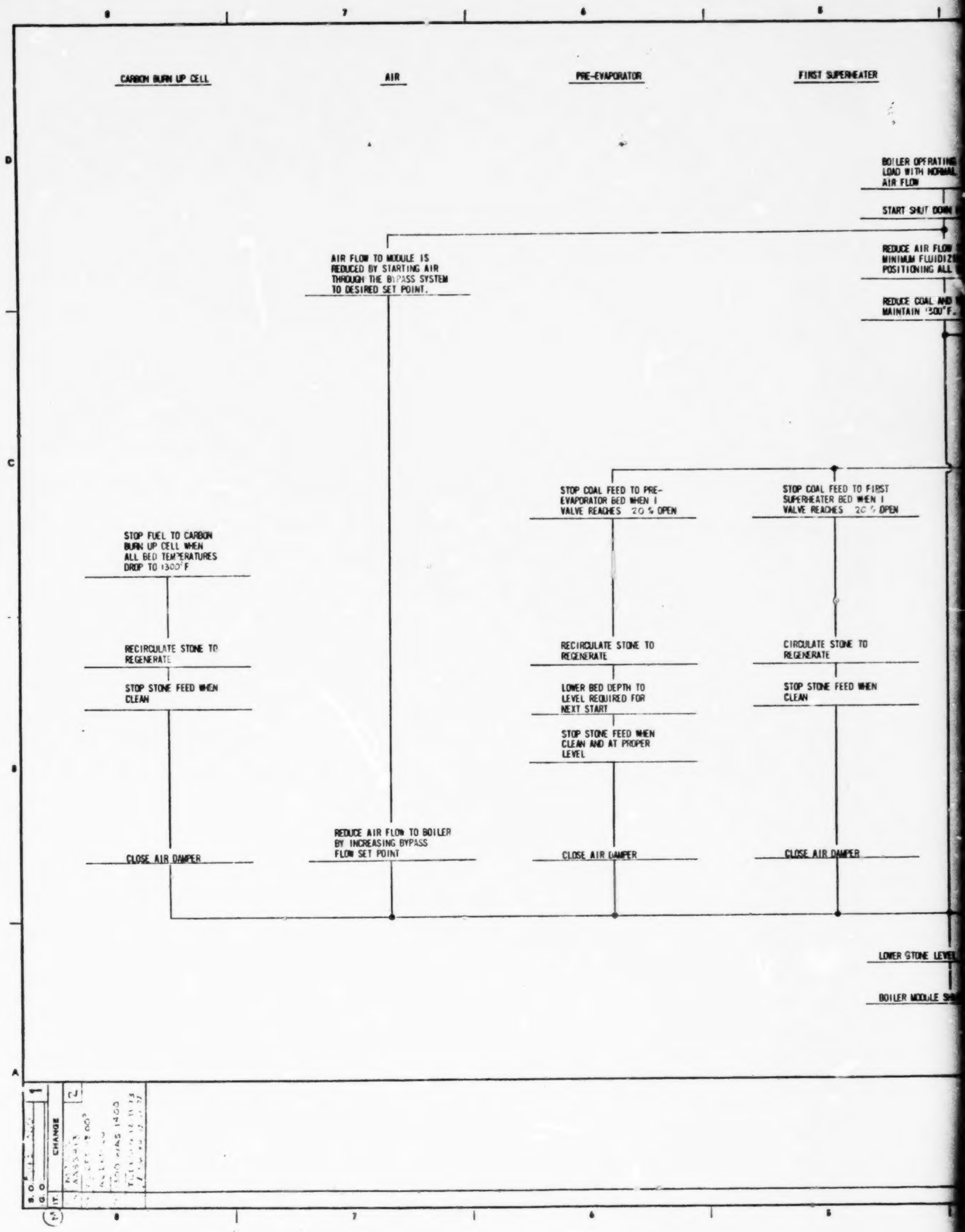
Figure 102 shows, in a simplified form, the steps required to shut down the first boiler module. The shutdown procedure is essentially parallel in the various parts of the plant.

The shutdown procedures that follow are based upon starting with the plant fully loaded and slowly decreasing the total plant load until all of the units have been shut down and the plant is off-line.

5.4.2 First Boiler Module Shutdown

The load will be reduced as described in section 5.3, on-line loading, by reducing the fuel input and water flow to a level that would permit one boiler module to be taken off the line.

Steam flow will be further reduced by positioning the airflow dampers to the boiler module and thereby permitting a reduction of water


8
7
6
5
CARBON BURN UP CELL
AIR
PRE-EVAPORATOR
FIRST SUPERHEATER
D
BOILER OPERATING
LOAD WITH NORMAL
AIR FLOW
START SHUT DOWN
AIR FLOW TO MODULE IS REDUCED BY STARTING AIR THROUGH THE BYPASS SYSTEM TO DESIRED SET POINT.
REDUCE AIR FLOW
MINIMUM FLUIDIZI
POSITIONING ALL
REDUCE COAL AND
MAINTAIN 1300°F.
C
STOP COAL FEED TO PRE-EVAPORATOR BED WHEN I VALVE REACHES 20 % OPEN
STOP COAL FEED TO FIRST SUPERHEATER BED WHEN I VALVE REACHES 20 % OPEN
STOP FUEL TO CARBON BURN UP CELL WHEN ALL BED TEMPERATURES DROP TO 1300°F
RECIRCULATE STONE TO REGENERATE
RECIRCULATE STONE TO REGENERATE
CIRCULATE STONE TO REGENERATE
STOP STONE FEED WHEN CLEAN
LOWER BED DEPTH TO LEVEL REQUIRED FOR NEXT START
STOP STONE FEED WHEN CLEAN
STOP STONE FEED WHEN CLEAN AND AT PROPER LEVEL
B
CLOSE AIR DAMPER
REDUCE AIR FLOW TO BOILER BY INCREASING BYPASS FLOW SET POINT
CLOSE AIR DAMPER
CLOSE AIR DAMPER
LOWER STONE LEVEL
BOILER MODULE SH
A
1
2
S. O.
G. O.
CHANGE
IT.
1300 WAS 1400
8
7
6
5

A

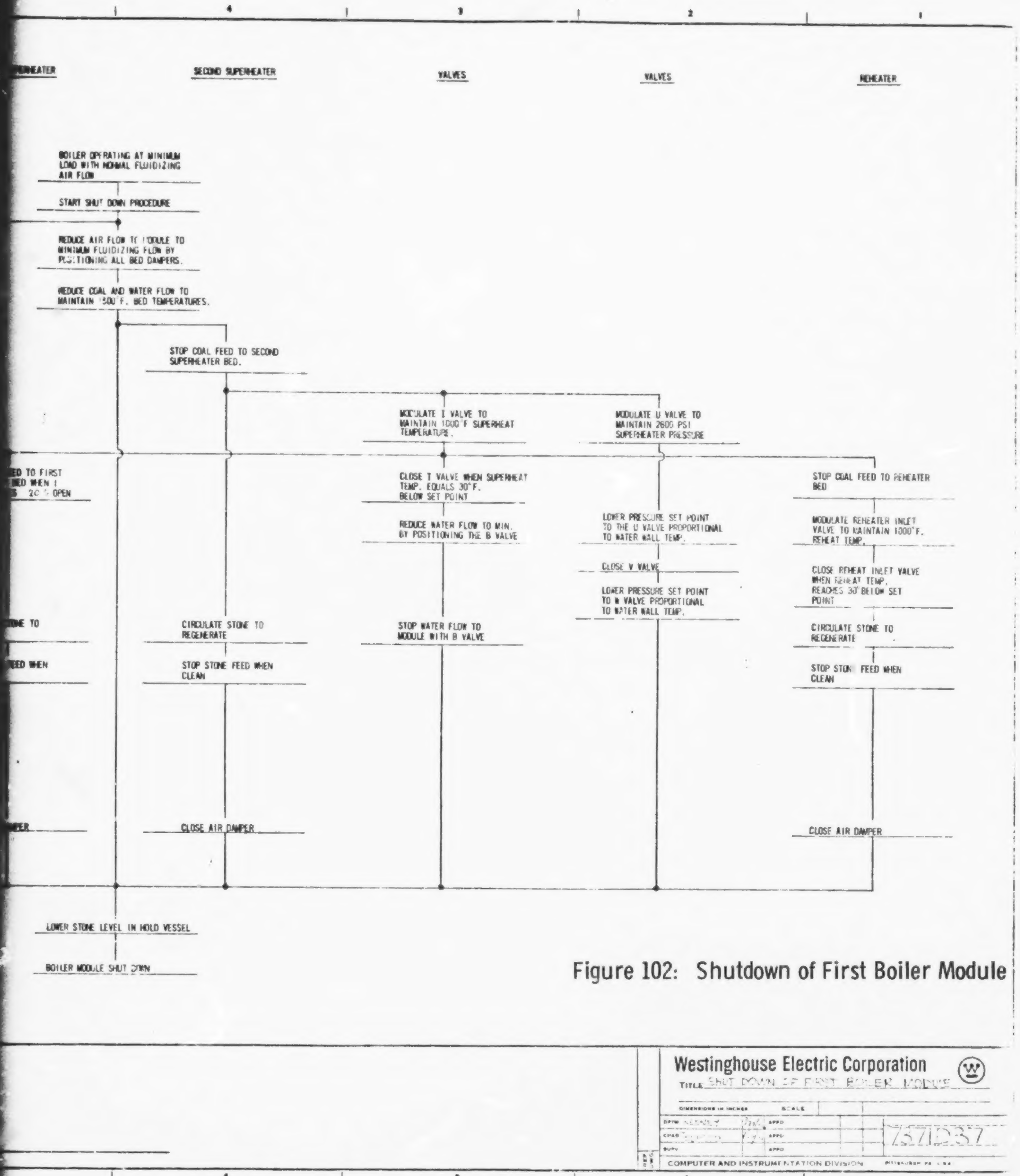

4
3
2
1
PERHEATER
SECOND SUPERHEATER
VALVES
VALVES
REHEATER
BOILER OPERATING AT MINIMUM LOAD WITH NORMAL FLUIDIZING AIR FLOW
START SHUT DOWN PROCEDURE
REDUCE AIR FLOW TO MODULE TO MINIMUM FLUIDIZING FLOW BY POSITIONING ALL BED DAMPERS.
REDUCE COAL AND WATER FLOW TO MAINTAIN 1500°F. BED TEMPERATURES.
STOP COAL FEED TO SECOND SUPERHEATER BED.
MODULATE I VALVE TO MAINTAIN 1000°F SUPERHEAT TEMPERATURE.
MODULATE U VALVE TO MAINTAIN 2600 PSI SUPERHEATER PRESSURE
ED TO FIRST
BED WHEN I
S 20 % OPEN
CLOSE T VALVE WHEN SUPERHEAT TEMP. EQUALS 30°F. BELOW SET POINT
STOP COAL FEED TO REHEATER BED
REDUCE WATER FLOW TO MIN. BY POSITIONING THE B VALVE
LOWER PRESSURE SET POINT TO THE U VALVE PROPORTIONAL TO WATER WALL TEMP.
MODULATE REHEATER INLET VALVE TO MAINTAIN 1000°F. REHEAT TEMP.
CLOSE V VALVE
CLOSE REHEAT INLET VALVE WHEN REHEAT TEMP. REACHES 30°BELOW SET POINT
LOWER PRESSURE SET POINT TO W VALVE PROPORTIONAL TO WATER WALL TEMP.
ONE TO
CIRCULATE STONE TO REGENERATE
STOP WATER FLOW TO MODULE WITH B VALVE
CIRCULATE STONE TO REGENERATE
EED WHEN
STOP STONE FEED WHEN CLEAN
STOP STONE FEED WHEN CLEAN
PER
CLOSE AIR DAMPER
CLOSE AIR DAMPER
LOWER STONE LEVEL IN HOLD VESSEL
BOILER MODULE SHUT DOWN
Westinghouse Electric Corporation
TITLE SHUT DOWN OF FIRST BOILER MODULE
DIMENSIONS IN INCHES
SCALE
DFTM
CHKD
SUPV
APPD
APPD
APPD
757D37
COMPUTER AND INSTRUMENTATION DIVISION
PITTSBURGH PA. U.S.A.

Figure 102: Shutdown of First Boiler Module

B

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flow and fuel flow. Total airflow to the pair of modules will be controlled by positioning the gas-turbine inlet guide vanes and bypassing air from the compressor outlet to the gas-turbine inlet. When the boiler module reaches its lowest controllable value, the module is ready to be shut down. Steam flow to the turbine will be further reduced by closing valve I and controlling the boiler superheater outlet pressure with the U valve. This requires adjusting the pressure set point down from the setting used for the relief valve function.

The coal flow to the second superheater will be stopped and the I valve closed enough to maintain a 1000°F superheater outlet temperature. This will reduce the heat in the bed and the amount and temperature of steam sent to the condenser. As the I valve closes, the pressure will be maintained by the U valve spilling steam to the condenser and by closing the B valve. A minimum temperature will be maintained in the preevaporator bed and the first-stage superheater bed. The I valve will close when the superheater outlet temperature reaches approximately 70°F below set point. This indicates that the maximum heat has been recovered from the secondary superheater bed. All of the steam from the boiler module will now be diverted to the condenser through the U valve, thus minimizing the steam-turbine temperature shock.

When the I valve starts to close, the coal feed to the reheater will be stopped and the reheater outlet steam temperature held at 1000°F by closing the reheater valve. This will permit the coal to be burned up in the reheater bed and the maximum amount of heat to be recovered. The steam will be directed to the other reheaters as the coal burns out. The reheat steam temperature will be held constant and the thermal shock to the steam turbine will be minimum.

The coal to the preevaporator bed and to the first superheater bed will be turned off when the I valve has been 80% closed. The pressure set point to the U valve will be lowered proportional to the water-wall outlet temperature, the V valve will close, and the W valve pressure set point will be adjusted downward as the water wall outlet temperature decreases because of the bed burn-out.

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After all of the coal has burned out of the beds, the stone will be recirculated until it has the correct chemical composition for the next start. It is assumed that this module would be the first of a pair to be restarted, so it will be necessary to reduce the levels of the preevaporator and first superheater beds. The stone recirculation equipment will then be shut down.

When the beds of the module have been restored to the conditions desired for the next start of the module, the air will be diverted by means of the gas-turbine by-pass damper and the boiler module air dampers. Water flow to the module will be stopped when the conditions are correct by closing the B valve.

The temperature control to the carbon burn-up cell will be adjusted so the fuel feed is stopped as the carbon discharge from the beds stops.

After all of the above have taken place, the boiler module is shut down but ready for the next start-up, when the load requirements demand it.

5.4.3 Second Boiler Module Shutdown

The description of the shutdown of the second boiler module will cover the shutdown of the second module of the first pair to be shut down. This shutdown will be followed by the shutdown of the first gas turbine. The second gas turbine with at least one boiler module and the steam turbine will still be in service.

The second boiler module will be shut down in the same manner as the first module except that none of the bed levels will be reduced and the gas turbine will be shut down. The bed levels will not be reduced since it is assumed that this will be the second module of a pair to be restarted.

The control system will maintain a minimum load of 4 MW on each gas-turbine generator. When one boiler module is not in service and the other has a minimum bed temperature of 1300°F, the gas-turbine inlet temperature will be approximately 1000°F. This temperature, with design airflow, will produce 4 MW. If the gas-turbine inlet temperature or the airflow decreases, the by-pass combustor must be ignited to maintain the minimum load. As the sections of the second boiler module are shut down, fuel to the by-pass combustor will be increased to maintain minimum load. When the bed stone has been regenerated for the next start-up, the gas turbine will be shut down and placed on turning gear ready for the next time the load requirements call for it to be started.

The booster fan and other auxiliary equipment will be shut down as the equipment is no longer needed.

5.4.4 Shutdown of Fourth Module

The shutdown of the fourth or last boiler module will require a different procedure because the steam temperature and reheat temperature cannot be maintained by the other boiler modules.

When the boiler module is at its lowest controllable level, the gas turbine should be at a minimum load and the steam turbine at approximately 10% of total load. The boiler module cannot be operated lower than this and still maintain bed temperature. The steam-turbine generator must be separated from the line and the governor valves closed quickly to shut down the boiler module. When this occurs, the fuel to all beds must be shut off and the boiler pressure controlled by the W and U valves. As in the case of the first boiler module start-up, the feedwater flow will now be controlled with the feed pump and the pressure controlled with the boiler valves.

The gas turbine may be kept in service to permit regeneration of the bed in preparation for the next start-up.

The shutdown of the steam turbine will follow the normal shutdown procedures for the steam plant. As the auxiliary equipment such as condensate pumps, boiler feed pump, etc., are no longer required, they will be shut down.

If no electrical load is required, the gas turbine can be placed on compressor discharge pressure control and the airflow controlled as required by varying the gas-turbine speed, positioning the gas-turbine compressor inlet guide vanes, and throttling the gas-turbine by-pass valve.

5.4.5 Emergency Shutdown

The plant will usually be shut down using the normal shutdown procedures described in the preceding discussion. Emergency conditions can occur, however, which will require emergency shutdown procedures. The emergency shutdown of a boiler module can result from several different types of failures. The plant trips are explained in Section 5.5. The major categories are as follows:

- Loss of airflow to the module caused by loss of the gas turbine or plugging of the distributor plate
- Failure in the water side of the module caused by failure of a pump or burn-out of bed tubes
- Loss of coal or stone feeding equipment
- Loss of other auxiliary equipment such as stone regeneration
- Steam-turbine trip as a result of loss of regenerator load
- Failure of particulate removal equipment causing particle overload
- Loss of load on gas-turbine generator.

5.4.5.1 Loss of Airflow

The loss of a gas turbine will result in the complete loss of airflow to two boiler modules. The plugging of the distributor will cause the loss of reduction of airflow to only one bed. If airflow is lost, the bed cannot be fluidized and the coal cannot be burned. The coal and stone feed will be stopped when the airflow stops. The steam temperature will drop because of the loss of heat input. When the superheater outlet temperature reaches 30°F below set point, the I valve will be closed. The U valve will be opened to maintain pressure in the module. The water flow will be reduced to minimum by throttling the B valve. The water wall-pressure will be lowered as the temperature decreases. The water wall flow will be stopped when the water outlet temperature equals the feedwater temperature. This indicates that the beds are cool.

The reheater valve will be closed when the reheater outlet is 30°F below set point.

The coal cannot be burned out, the stone regenerated, or the preevaporator bed lowered because the beds are not fluidized. The beds will be emptied when cold and prepared for the next start-up.

5.4.5.2 Failure in the Water Side

Water side failures can be caused by an external problem such as failure of a boiler feed pump, feedwater heater, condenser, or other auxiliaries; or by an internal failure such as a tube rupture.

Failure of the feedwater supply will result in the tripping of the module to prevent burning out the tubes.

Tube rupture can be severe or minor, depending on the number of tubes and the size of the failure. If the rupture is small, the module will be shut down in the normal manner. However, a large failure can result in an increase in bed pressure or carry-over of the bed caused by jetting. If the emergency shutdown is required, the I valve and B valve will be closed and the U valve opened to release water and steam to the condenser. The coal feed will be stopped and the bed left in operation to burn out the coal, regenerate the stone, and lower the bed so repairs can be made. The reheater will be shut down in the normal manner if the rupture is not in the reheater. If the problem is in the reheater, however, the reheater inlet valve will be closed and the coal stopped to the reheater bed only. The reheat steam flow will be diverted to the other reheaters and the remainder of the module shut down in the normal manner.

5.4.5.3 Loss of Coal or Stone Feeding Equipment

If the coal feeding equipment to the reheater or to the final superheater beds fails, the boiler module can be shut down in the normal manner because these are the first coal feeds to be turned off. If the failure is to the preevaporator or first superheater bed, all of the coal feed must be stopped at the same time because the water wall or first superheater temperature will drop and prevent normal boiler operation.

If the stone feeding equipment malfunctions, the module will be shut down in the normal manner, except that the beds will not be regenerated.

5.4.5.4 Loss of Auxiliary Equipment

The loss of auxiliary or other associated equipment can result in a normal shutdown, a load reduction, or an emergency shutdown. All of these conditions cannot be determined until the plant design is completed.

5.4.5.5 Steam-Turbine Trip

A steam-turbine trip will call for a trip or emergency shutdown of all of the boiler modules. This emergency will require the same procedure as described above for a water-side problem.

5.4.5.6 Failure of Particulate Removal Equipment

The failure of the particulate removal equipment will require removing the associated gas turbine from the line to prevent erosion of the turbine blades. The boiler modules associated with this gas turbine will also have to be shut down and, if this is the only gas turbine on the line, the steam turbine will have to be shut down because no steam will be available.

5.4.6 Dolomite Sorbent Regeneration System Shutdown

Since the sorbent regeneration system requires a supply of CO_2 in order to operate, it will normally be shut down before the last boiler module. Therefore, the stone can be clean in only three of the four boiler modules. However, it may be possible to obtain sufficient CO_2 when the gas turbine is operating on supplemental fuel oil. In this event, all of the boiler modules can be shut down with clean sorbent.

Stone recirculation continues until after the $CaSO_4$ reducer vessel and the H_2S generator vessel have been shut down. The $CaSO_4$ reducer vessel will be shut down by stopping the feed of the coal, air, and steam to the gas producer.

As the vessel pressure drops, the expander throttle valve from the gas-turbine air compressor to the expander will be opened to maintain CO_2 flow to the H_2S generator. The exit gas throttle valve will be closed so all of the gas will be sent to the stack through the by-pass valve.

The CO_2 scrubber stripper flow will be stopped, the cooling water to the condenser stopped, the spray flow stopped, and the bleed valve opened as needed to prevent surge on the compressor. The expander will be stopped by closing the throttle valve.

The stone feed will be stopped when the stone has been transferred from the hold vessels to the surge vessel so the stone will be in the proper place for a start-up.

The last boiler module must be shut down after the stone feed has stopped.

5.5 TRIPS

A complex power plant requires a plant trip scheme to interlock the various plant components. Figure 103, trip chart, shows the effect on each of the plant components if one plant component is tripped. The vertical columns represent the contingency condition and the horizontal rows, the effect. For example, if the no. 1 gas turbine trips, the no. 1 and no. 2 boiler modules must be tripped with no effect on the no. 3 and no. 4 boiler modules or the no. 2 gas turbine, and so on.

	GAS TURBINE NO. 1 TRIP	GAS TURBINE GEN. NO. 1 TRIP	BOILER MODULE NO. 1 TRIP	BOILER MODULE NO. 2 TRIP	GAS TURBINE NO. 2 TRIP	GAS TURBINE GEN. NO. 2 TRIP
GAS TURBINE NO. 1	SHUT DOWN	1. IDLE SPEED 2. CONTROL DISCH. PRESS. 3. CONTROL AIR FLOW 4. BLEED HOT GAS	IF FIRST – 1. CONTROL AIR FLOW IF SECOND – 1. SHUT DOWN	IF FIRST 1. CONTROL AIR FLOW IF SECOND 1. SHUT DOWN		
BOILER MODULE NO. 1	TRIP	1. GO TO MIN LOAD	1. CLOSE I 2. CONTROL PRESS – U 3. SET FLOW – B 4. BURN OUT FUEL OR SHUT OFF AIR	———		
BOILER MODULE NO. 2	TRIP	1. GO TO MIN LOAD	———	1. CLOSE I 2. CONTROL PRESS – V 3. SET FLOW – B 4. BURN OUT FUEL OR SHUT OFF AIR		
GAS TURBINE NO. 2	———	———	———	———		
BOILER MODULE NO. 3	———	———	———	———		
BOILER MODULE NO. 4	———	———	———	———		
STEAM TURBINE	1. LOAD IS REDUCED	1. LOAD IS REDUCED	1. IF LAST MODULE TRIP 2. REDUCE LOAD	1. IF LAST MODULE TRIP 2. REDUCE LOAD		
STONE FEED	1. IF FIRST – OK 2. IF SECOND – STOP	———	1. IF LAST SHUT DOWN	1. IF LAST SHUT DOWN		
CA SO_4 REDUCER	1. IF FIRST – OK 2. IF SECOND – ISOLATE	———	1. IF LAST SHUT DOWN	1. IF LAST SHUT DOWN		
H_2 S GENERATOR	1. IF FIRST – OK 2. IF SECOND – ISOLATE	———	1. IF LAST SHUT DOWN	1. IF LAST SHUT DOWN		

S.O. 1

IT. 2

CHANGE

A

GAS TURBINE GEN. NO. 2	BOILER MODULE NO. 3	BOILER MODULE NO. 4	STEAM TURBINE	STONE FEED	Ca SO_4 REDUCER	H_2S GENERATOR
TRIP	TRIP	TRIP	TRIP			
			1. SHUT DOWN AFTER STONE CLEAN 2. GO AUX FUEL 3. IDLE SPEED	—	—	—
			1. STOP FUEL FEED 2. CLOSE I OPEN 3. SHUT DOWN WHEN STONE CLEAN	1. REDUCE LOAD 2. SHUT DOWN	GO TO ONCE THROUGH	GO TO ONCE THROUGH
			1. STOP FUEL FEED 2. CLOSE I OPEN 3. SHUT DOWN WHEN STONE CLEAN	1. REDUCE LOAD 2. SHUT DOWN	GO TO ONCE THROUGH	GO TO ONCE THROUGH
			—	—	—	—
			—	1. REDUCE LOAD 2. SHUT DOWN	GO TO ONCE THROUGH	GO TO ONCE THROUGH
			—	1. REDUCE LOAD 2. SHUT DOWN	GO TO ONCE THROUGH	GO TO ONCE THROUGH
			SHUT DOWN	—	—	—
			1. CONTINUE IF POSSIBLE TO CLEAN STONE 2. STOP WHEN STONE IS CLEAN	—	GO TO ONCE THROUGH	GO TO ONCE THROUGH
			1. STOP WHEN STONE IS CLEAN	SHUT DOWN WHEN HOLD VES. EMPTY OR SURGE FULL	—	SHUT DOWN
			1. STOP WHEN STONE IS CLEAN	SHUT DOWN WHEN HOLD VES. EMPTY OR SURGE FULL	SHUT DOWN	—
			IF GEN. ONLY KEEP 1 MOD OPERATING TO SUPPLY STEAM			

Figure 103: Trip Chart

Westinghouse Electric Corporation

TITLE TRIP CHART

DIMENSIONS IN INCHES SCALE

DFTR

CHKD

SUPV

APPD

APPD

APPD

COMPUTER AND INSTRUMENTATION DIVISION

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5.6 RUNBACKS, RUN-UPS, AND LIMITS

A complex plant requires a system of runbacks, run-ups, and limits to maintain plant load at a safe value in case something unforeseen occurs. This system will override load demand to adjust the operating level as required by such a contingency. The system must contain three parts; they are defined as follows:

- Runback -- a reduction of load initiated by an alarm condition. The load demand will be reduced until the alarm clears. The runback can occur at any load level and reduce the load demand to any value required to clear the alarm condition. Minimum limits cannot be exceeded in clearing the fault. The control items that will be included in the runback scheme are low excess air, high bed temperatures, insufficient feedwater flow with both pumps in operation, approaching the surge condition on a gas turbine, and others.
- Run-up -- an increase of load initiated by a low alarm. The load demand will be increased as much as needed to clear the alarm. The operating value at which this condition can occur varies with the fault and cannot be defined as an absolute value. Maximum limits cannot be exceeded to clear faults. Runbacks or high limits will always have precedence over run-ups because it is usually safer to reduce the load than to increase it. If necessary, a portion of the plant will be tripped to clear the alarm condition. Examples of run-ups are low bed temperature, minimum water flow to a boiler module, minimum load on a gas-turbine generator, and others.
- Limits -- high or low operating restrictions placed on the plant based on the equipment in service. Limits are different from runbacks and run-ups in that they have absolute known values. If a high limit exists and the plant is operating below that value, it does not affect operation as long as

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the load demand does not reach that value. The load demand cannot be increased above the value of the limit. If the value exceeds the limit value at any time, the load will be reduced to the limit value as rapidly as possible. Control system limits are imposed by critical equipment such as pumps, fans, and others. These pieces of equipment are essential for safe plant operation above some value, as, for example, with two 60% capacity boiler feed pumps: the plant cannot operate satisfactorily and safely above approximately 55% of maximum capacity with only one pump in service.

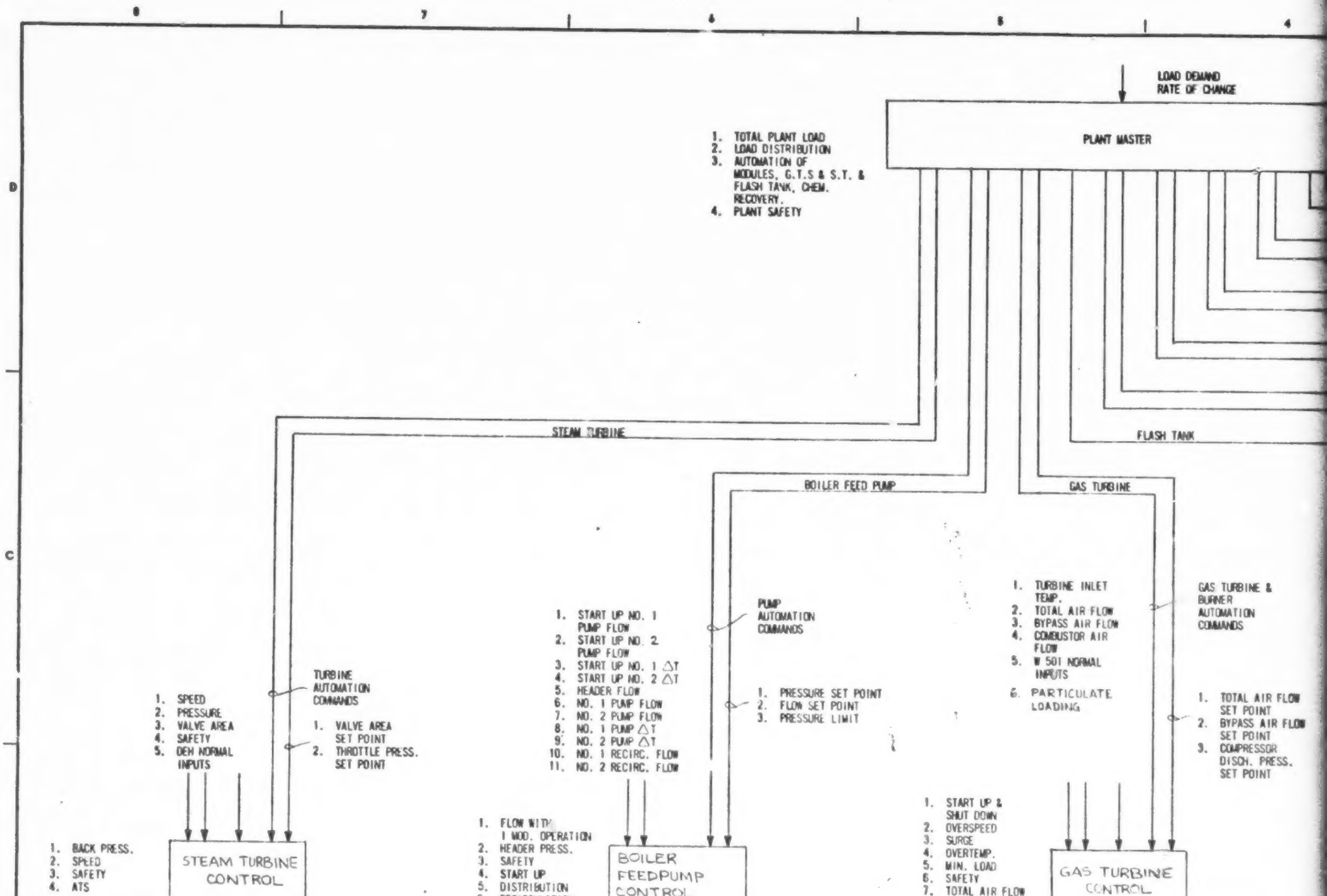
The runback, run-up, and limit system will be a part of the master control and will be operational only when the plant is on automatic control and fully coordinated. When operating in manual or not coordinated, the alarm system will warn the operator of contingency conditions, and, if necessary, the trip system will trip part or all of the plant to prevent damage.

5.7 TOTAL PLANT CONTROL SYSTEM

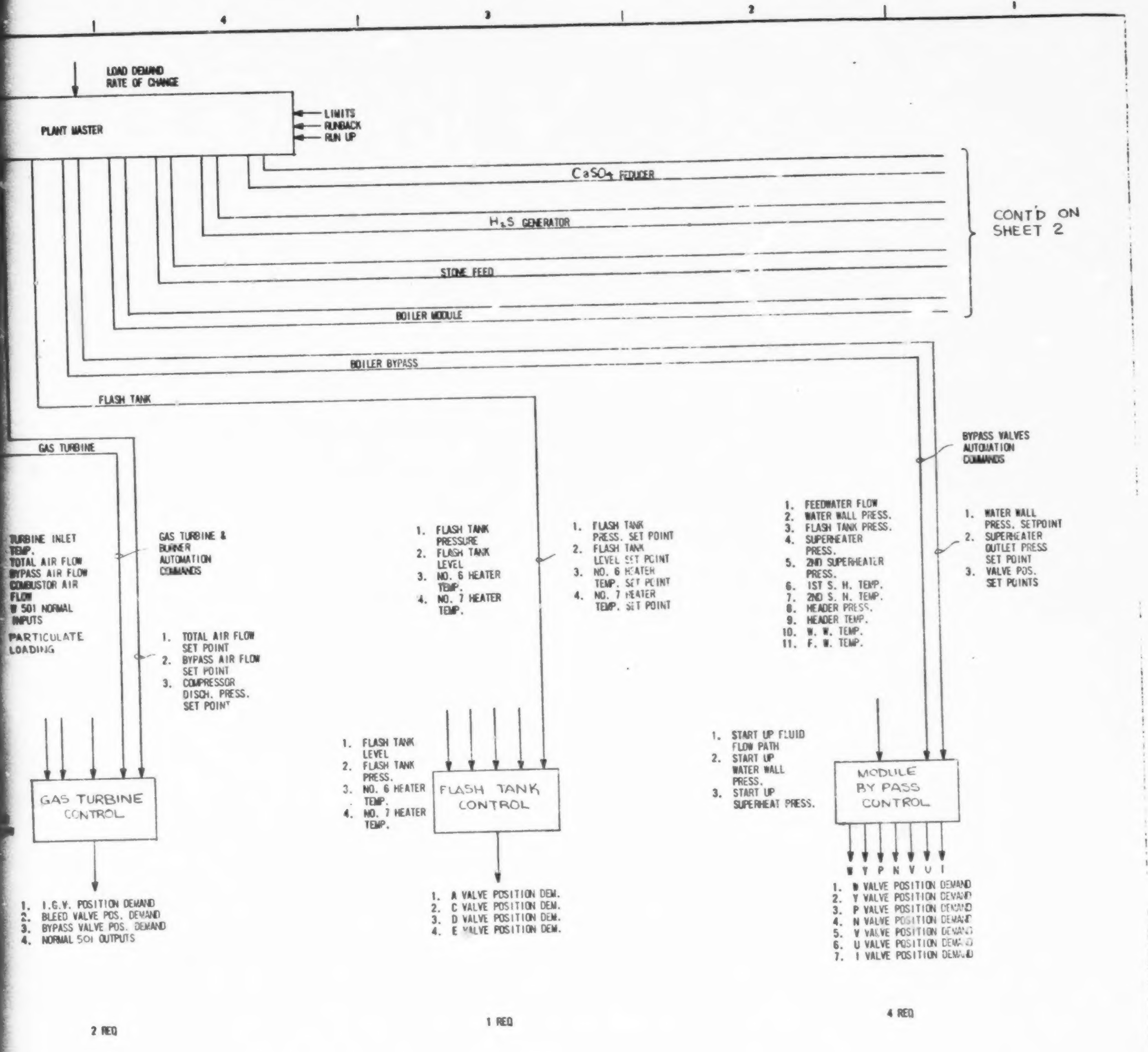
5.7.1 Introduction

The total plant control system will be a hybrid, integrated system made up of a combination of digital and analog control and information equipment. The control functions that will be performed by this system are shown on Figure 104. This drawing shows ten systems, each designated by a block on the drawing. The inputs and outputs to each system, as well as the control functions performed within that system, are shown on the drawing.

A total analog control system, with a digital information system of the type currently used on conventional power plants, could be applied to the fluidized bed boiler control; but the advantages of the digital system for large automated and adaptive control applications could not be incorporated. The fluidized bed boiler will lend itself to control


LOAD DEMAND
RATE OF CHANGE
PLANT MASTER
1. TOTAL PLANT LOAD
2. LOAD DISTRIBUTION
3. AUTOMATION OF MODULES, G.T.S & S.T. & FLASH TANK, CHEM. RECOVERY.
4. PLANT SAFETY
STEAM TURBINE
BOILER FEED PUMP
GAS TURBINE
FLASH TANK
1. SPEED
2. PRESSURE
3. VALVE AREA
4. SAFETY
5. DEH NORMAL INPUTS
TURBINE AUTOMATION COMMANDS
1. VALVE AREA SET POINT
2. THROTTLE PRESS. SET POINT
1. BACK PRESS.
2. SPEED
3. SAFETY
4. ATS
STEAM TURBINE CONTROL
NORMAL DEH OUTPUTS
1 REQ
1. START UP NO. 1 PUMP FLOW
2. START UP NO. 2 PUMP FLOW
3. START UP NO. 1 △T
4. START UP NO. 2 △T
5. HEADER FLOW
6. NO. 1 PUMP FLOW
7. NO. 2 PUMP FLOW
8. NO. 1 PUMP △T
9. NO. 2 PUMP △T
10. NO. 1 RECIRC. FLOW
11. NO. 2 RECIRC. FLOW
PUMP AUTOMATION COMMANDS
1. PRESSURE SET POINT
2. FLOW SET POINT
3. PRESSURE LIMIT
1. FLOW WITH 1 MOD. OPERATION
2. HEADER PRESS.
3. SAFETY
4. START UP
5. DISTRIBUTION
6. RECIRCULATION
7. PRESS. LIMIT FOR START UP
BOILER FEEDPUMP CONTROL
1. NO. 1 PUMP SPEED DEM
2. NO. 2 PUMP SPEED DEM
3. NO. 1 RECIRC VALVE POS DEM
4. NO. 2 RECIRC VALVE POS DEM
5. NO. 1 START UP VALVE POS DEM
6. NO. 2 START UP VALVE POS DEM
7. NO. 1 START UP RECIRC VALVE POS DEM
8. NO. 2 START UP RECIRC VALVE POS DEM
1 REQ
1. TURBINE INLET TEMP.
2. TOTAL AIR FLOW
3. BYPASS AIR FLOW
4. COMBUSTOR AIR FLOW
5. W 501 NORMAL INPUTS
6. PARTICULATE LOADING
GAS TURBINE & BURNER AUTOMATION COMMANDS
1. TOTAL AIR FLOW SET POINT
2. BYPASS AIR FLOW SET POINT
3. COMPRESSOR DISCH. PRESS. SET POINT
1. START UP & SHUT DOWN
2. OVERSPEED
3. SURGE
4. OVERTEMP.
5. MIN. LOAD
6. SAFETY
7. TOTAL AIR FLOW
8. BYPASS AIR FLOW
9. COMPRESSOR DISCH. PRESS.
10. SAFETY
GAS TURBINE CONTROL
1. I.G.V. POSITION DEMAND
2. BLEED VALVE POS. DEMAND
3. BYPASS VALVE POS. DEMAND
4. NORMAL 501 OUTPUTS
2 REQ
8
7
6
5
4
D
C
B
A
CHANGE

A



4
3
2
1
LOAD DEMAND
RATE OF CHANGE
PLANT MASTER
LIMITS
RUNBACK
RUN UP
CaSO4 REDUCER
H2S GENERATOR
STONE FEED
BOILER MODULE
CONT'D ON SHEET 2
BOILER BYPASS
FLASH TANK
GAS TURBINE
TURBINE INLET TEMP.
TOTAL AIR FLOW
BYPASS AIR FLOW
COMBUSTOR AIR FLOW
W 501 NORMAL INPUTS
PARTICULATE LOADING
GAS TURBINE & BURNER AUTOMATION COMMANDS
1. TOTAL AIR FLOW SET POINT
2. BYPASS AIR FLOW SET POINT
3. COMPRESSOR DISCH. PRESS. SET POINT
GAS TURBINE CONTROL
1. I.G.V. POSITION DEMAND
2. BLEED VALVE POS. DEMAND
3. BYPASS VALVE POS. DEMAND
4. NORMAL 501 OUTPUTS
2 REQ
1. FLASH TANK PRESSURE
2. FLASH TANK LEVEL
3. NO. 6 HEATER TEMP.
4. NO. 7 HEATER TEMP.
1. FLASH TANK PRESS. SET POINT
2. FLASH TANK LEVEL SET POINT
3. NO. 6 HEATER TEMP. SET POINT
4. NO. 7 HEATER TEMP. SET POINT
1. FLASH TANK LEVEL
2. FLASH TANK PRESS.
3. NO. 6 HEATER TEMP.
4. NO. 7 HEATER TEMP.
FLASH TANK CONTROL
1. A VALVE POSITION DEM.
2. C VALVE POSITION DEM.
3. D VALVE POSITION DEM.
4. E VALVE POSITION DEM.
1 REQ
BYPASS VALVES AUTOMATION COMMANDS
1. FEEDWATER FLOW
2. WATER WALL PRESS.
3. FLASH TANK PRESS.
4. SUPERHEATER PRESS.
5. 2ND SUPERHEATER PRESS.
6. 1ST S. H. TEMP.
7. 2ND S. H. TEMP.
8. HEADER PRESS.
9. HEADER TEMP.
10. W. W. TEMP.
11. F. W. TEMP.
1. WATER WALL PRESS. SETPOINT
2. SUPERHEATER OUTLET PRESS SET POINT
3. VALVE POS. SET POINTS
1. START UP FLUID FLOW PATH
2. START UP WATER WALL PRESS.
3. START UP SUPERHEAT PRESS.
MODULE BY PASS CONTROL
W Y P N V U I
1. W VALVE POSITION DEMAND
2. Y VALVE POSITION DEMAND
3. P VALVE POSITION DEMAND
4. N VALVE POSITION DEMAND
5. V VALVE POSITION DEMAND
6. U VALVE POSITION DEMAND
7. I VALVE POSITION DEMAND
4 REQ

Figure 104A: Control System Functions

Westinghouse Electric Corporation
TITLE CONTROL SYSTEM FUNCTIONS
DIMENSIONS IN INCHES SCALE
DFTM [illegible] APPD
CHKD [illegible] APPD
SUPV APPD
7371D42
SHEET 1 of 2
REF DWG
COMPUTER AND INSTRUMENTATION DIVISION
PITTSBURGH PA. U.S.A.

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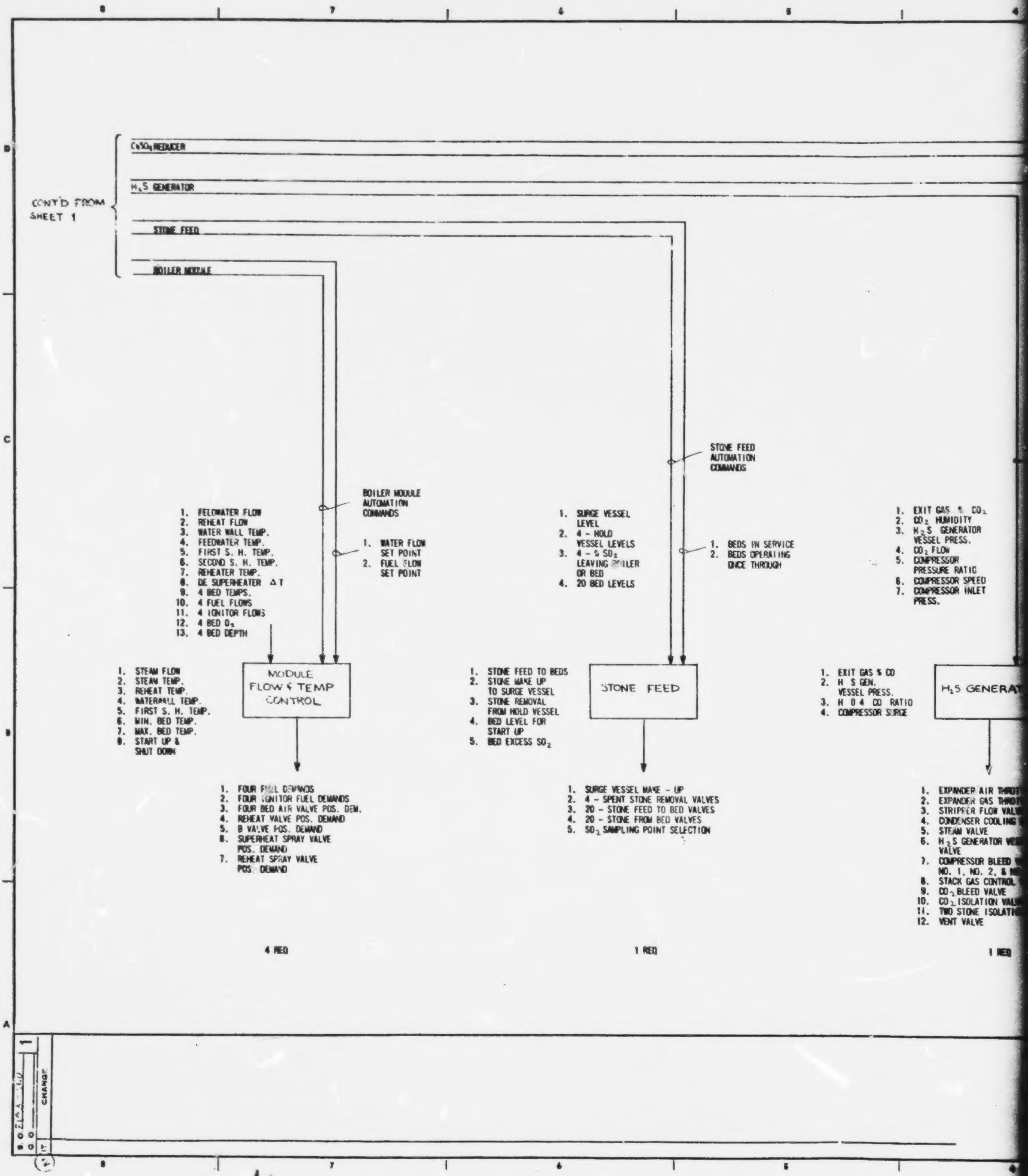
8
7
6
5
4
D
C
B
A
CONT'D FROM SHEET 1
$CaSO_4$ REDUCER
H_2S GENERATOR
STONE FEED
BOILER MODULE
BOILER MODULE AUTOMATION COMMANDS
1. WATER FLOW SET POINT
2. FUEL FLOW SET POINT
1. FEEDWATER FLOW
2. REHEAT FLOW
3. WATER WALL TEMP.
4. FEEDWATER TEMP.
5. FIRST S. H. TEMP.
6. SECOND S. H. TEMP.
7. REHEATER TEMP.
8. DE SUPERHEATER Δ T
9. 4 BED TEMPS.
10. 4 FUEL FLOWS
11. 4 IGNITOR FLOWS
12. 4 BED O_2
13. 4 BED DEPTH
1. STEAM FLOW
2. STEAM TEMP.
3. REHEAT TEMP.
4. WATERWALL TEMP.
5. FIRST S. H. TEMP.
6. MIN. BED TEMP.
7. MAX. BED TEMP.
8. START UP & SHUT DOWN
MODULE FLOW & TEMP CONTROL
1. FOUR FUEL DEMANDS
2. FOUR IGNITOR FUEL DEMANDS
3. FOUR BED AIR VALVE POS. DEM.
4. REHEAT VALVE POS. DEMAND
5. B VALVE POS. DEMAND
6. SUPERHEAT SPRAY VALVE POS. DEMAND
7. REHEAT SPRAY VALVE POS. DEMAND
4 REQ
STONE FEED AUTOMATION COMMANDS
1. SURGE VESSEL LEVEL
2. 4 - HOLD VESSEL LEVELS
3. 4 - % SO_2 LEAVING BOILER OR BED
4. 20 BED LEVELS
1. BEDS IN SERVICE
2. BEDS OPERATING ONCE THROUGH
1. STONE FEED TO BEDS
2. STONE MAKE UP TO SURGE VESSEL
3. STONE REMOVAL FROM HOLD VESSEL
4. BED LEVEL FOR START UP
5. BED EXCESS SO_2
STONE FEED
1. SURGE VESSEL MAKE - UP
2. 4 - SPENT STONE REMOVAL VALVES
3. 20 - STONE FEED TO BED VALVES
4. 20 - STONE FROM BED VALVES
5. SO_2 SAMPLING POINT SELECTION
1 REQ
1. EXIT GAS % CO_2
2. CO_2 HUMIDITY
3. H_2S GENERATOR VESSEL PRESS.
4. CO_2 FLOW
5. COMPRESSOR PRESSURE RATIC
6. COMPRESSOR SPEED
7. COMPRESSOR INLET PRESS.
1. EXIT GAS % CO
2. H S GEN. VESSEL PRESS.
3. H 0 4 CO RATIO
4. COMPRESSOR SURGE
H_2S GENERAT
1. EXPANDER AIR THROT
2. EXPANDER GAS THROT
3. STRIPPER FLOW VALV
4. CONDENSER COOLING
5. STEAM VALVE
6. H_2S GENERATOR VE
VALVE
7. COMPRESSOR BLEED
NO. 1, NO. 2, &
8. STACK GAS CONTROL
9. CO_2 BLEED VALVE
10. CO_2 ISOLATION VAL
11. TWO STONE ISOLATI
12. VENT VALVE
1 REQ
1
CHANGE
LT

A

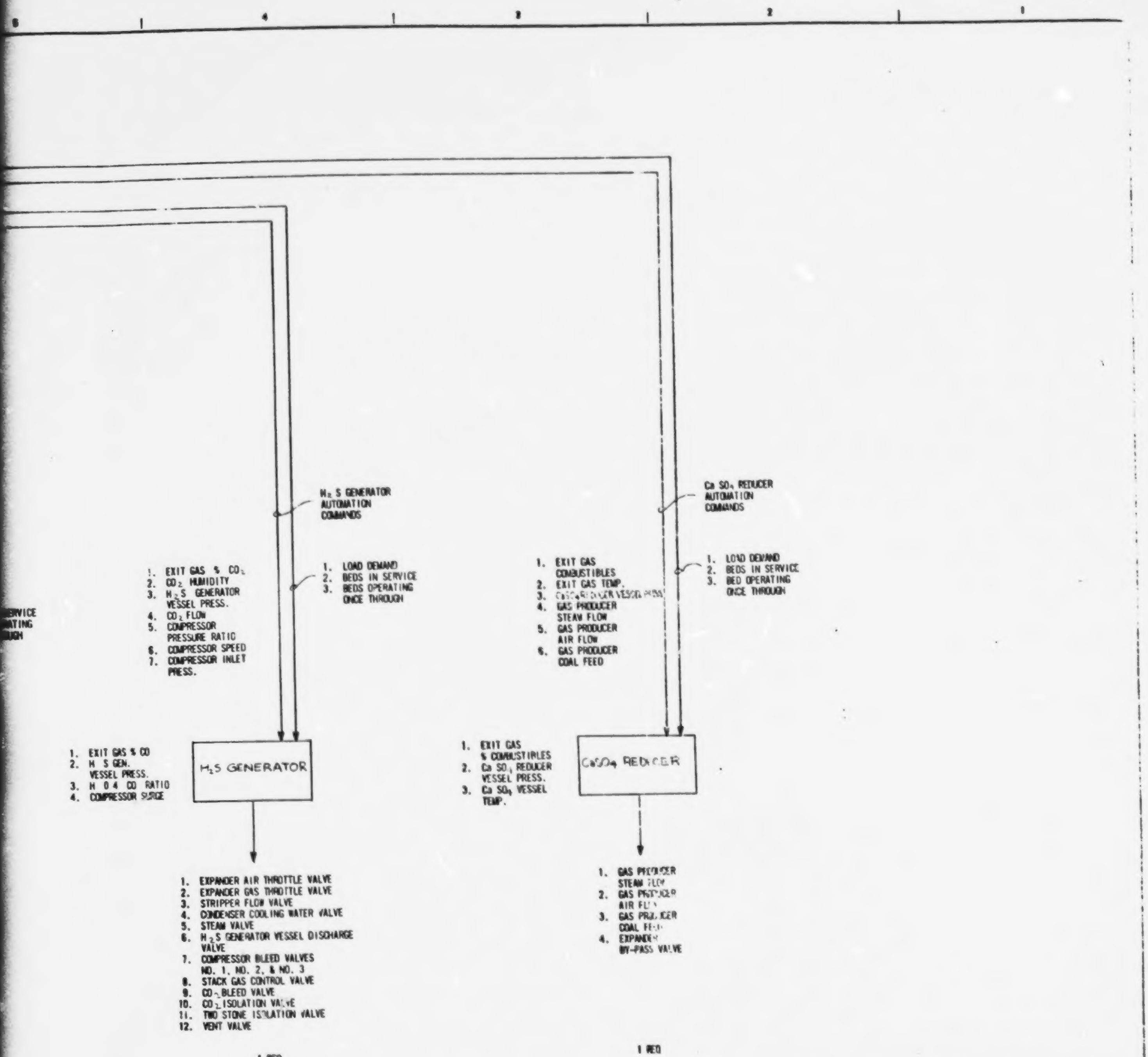
H₂ S GENERATOR AUTOMATION COMMANDS
1. LOAD DEMAND 2. BEDS IN SERVICE 3. BEDS OPERATING ONCE THROUGH
1. EXIT GAS % CO₂ 2. CO₂ HUMIDITY 3. H₂S GENERATOR VESSEL PRESS. 4. CO₂ FLOW 5. COMPRESSOR PRESSURE RATIO 6. COMPRESSOR SPEED 7. COMPRESSOR INLET PRESS.
1. EXIT GAS % CO 2. H S GEN. VESSEL PRESS. 3. H O 4 CO RATIO 4. COMPRESSOR SURGE
H₂S GENERATOR
1. EXPANDER AIR THROTTLE VALVE 2. EXPANDER GAS THROTTLE VALVE 3. STRIPPER FLOW VALVE 4. CONDENSER COOLING WATER VALVE 5. STEAM VALVE 6. H₂S GENERATOR VESSEL DISCHARGE VALVE 7. COMPRESSOR BLEED VALVES NO. 1, NO. 2, & NO. 3 8. STACK GAS CONTROL VALVE 9. CO₂ BLEED VALVE 10. CO₂ ISOLATION VALVE 11. TWO STONE ISOLATION VALVE 12. VENT VALVE
1 REQ
Ca SO₄ REDUCER AUTOMATION COMMANDS
1. LOAD DEMAND 2. BEDS IN SERVICE 3. BED OPERATING ONCE THROUGH
1. EXIT GAS COMBUSTIBLES 2. EXIT GAS TEMP. 3. CaSO₄ REDUCER VESSEL PRESS. 4. GAS PRODUCER STEAM FLOW 5. GAS PRODUCER AIR FLOW 6. GAS PRODUCER COAL FEED
1. EXIT GAS % COMBUSTIBLES 2. Ca SO₄ REDUCER VESSEL PRESS. 3. Ca SO₄ VESSEL TEMP.
CaSO₄ REDUCER
1. GAS PRODUCER STEAM FLOW 2. GAS PRODUCER AIR FLOW 3. GAS PRODUCER COAL FEED 4. EXPANDER BY-PASS VALVE
1 REQ
SERVICE
ATING
UGH

Figure 104B: Control System Functions

Westinghouse Electric Corporation
TITLE CONTROL SYSTEM FUNCTIONS
DIMENSIONS IN INCHES SCALE
DRWN APPD
CHKD APPD
SUPV APPD
737D42
SHEET 2
COMPUTER AND INSTRUMENTATION DIVISION PITTSBURGH PA. U.S.A.

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optimization more readily than the conventional boiler because of the ability to control each module or portion of the boiler independently.

Analog control equipment will be used in certain critical control loops and in the safety systems. Digital control equipment will be used for plant automation, plant coordination, and adaptive control. Analog and digital equipment will be combined in the conventional boiler and turbine control loops. The total system will be arranged to utilize the advantages of the two types of control techniques and will be arranged to permit total plant automatic operation, system automatic operation, and manual operation.

The following section briefly describes the function performed by the ten control systems.

5.7.2 Plant Master Control System

All complex power plants need a master control system to direct the control systems for the various plant components, to provide central automation, to accept demand signals from the load dispatcher, and to provide a common point for operator interface.

The master system controls total plant load from a remote set point or an operator set value. It distributes the load to the various plant components to provide the desired type of operation as described above.

It will also provide the automation commands to start or stop plant components as required to meet the desired load. Plant contingency situations will be handled from the master with the necessary action commands sent to the control systems and subsystems.

A system of panel interface with permissive and reject logic will be provided to permit the operator to select safe operating modes and configurations.

The master system will consist mostly of control logic systems rather than feedback systems.

The input signals to the master will be plant load and load rate of change from the dispatcher or operator, limits from the plant components or the operator, runback or run-up demands from the plant components, mode selection requests from the operator, and operating status information from the operator.

The output signals will be the various control set points to the plant systems and subsystems, and automation commands to the plant components.

5.7.3 Steam Turbine

A conventional single reheat steam turbine utilizing an electrohydraulic control system and automatic turbine start-up will be employed.

This control system is different from that of the standard digital electrohydraulic system because the plant is operated in the turbine-following rather than the usual boiler-following manner. In the turbine-following system, the plant load is changed by adjusting fuel input, and the turbine inlet pressure is controlled by adjusting the turbine governor valves. In the boiler-following system, the plant load is changed by adjusting the governor valves, and the turbine inlet pressure is controlled by adjusting the fuel input. The on-line control system will be of the feed-forward type, with the master control requesting valve area and this signal trimmed to maintain desired throttle pressure. The safety systems will be the same as for a conventional turbine except for the throttle pressure control which is included in the base control system. Valve management, valve testing, overspeed protection control, and speed error controls will be included.

The calculated rotor stress will provide a maximum limit for the rate of change of load as permitted by the turbine metal temperatures at that instant.

The turbine acceleration program will contain all of the features of the standard programs and is not described further here.

The pressure ramping of the first boiler module will be done with floating governor valves position rather than fixed valve position. This will accommodate the turbine-following control approach and make automation at the start-up easier, as the valves will always be in the proper position at the end of the ramp.

The speed error can be limited in the increasing direction so the plant can be operated in a base loaded manner without frequency participation. The speed control will function only as overspeed protection during this mode of operation.

The analog and digital inputs and outputs to the steam-turbine control system will be the same as for a standard 500 MW single reheat machine except that the master control will provide a demand for valve area (PI/PT) and the throttle pressure set point. Also, mode selection logic can be initiated by the master control as well as by the operator.

5.7.4 Boiler Feed Pumps

The boiler feed pump control system will perform several functions in addition to those of the electrohydraulic speed and the safety control provided with the pumps:

- Feedwater flow control during the first boiler module start-up
- Maximum pressure override during the first boiler module start-up
- Feedwater header pressure control after the first boiler module is off the flash tank
- Feedwater pump flow balancing when two pumps are in operation
- Feedpump recirculation control
- Automatic start-up and shutdown of the feedpumps

The above listed control circuits are standard and will not be described here.

The master control will provide set points for flow and pressure control plus automatic signals requesting start-up and shutdown of the feedpumps. Mode selection logic can be initiated by the master control as well as by the operator.

The input signals to the feedpump control, in addition to those provided with the electrohydraulic control system, will be feedwater header pressure, individual feedpump flow, individual feedpump temperature rise and recirculation flow. The output signals will be individual feedpump speed demand and individual feedpump recirculation valve position demand. The logic signals provided will be those required by the turbine pump manufacturers for automation.

The input signals for the start-up feedpumps will be the individual pump flows and temperature rise. The outputs from the start-up pump controls will be control valve position demand and recirculation valve position demand.

5.7.5 Gas Turbine

The gas turbine and gas-turbine control system will be different from a W501 in the following areas:

- A separate combustion chamber will be used.
- The fuel system will be smaller in capacity because only minimum load fuel will be required.
- Air can flow through the boiler modules, the by-pass combustor, or both.
- Inlet guide vanes will be used to control airflow.
- The by-pass combustor will not be needed after the first boiler module has reached approximately minimum load.
- Turbine inlet temperatures can be measured.
- Compressor airflow will be measured.
- Peak temperature limits will not be required because of the low (1750°F) maximum bed temperature.

- After start-up, only minimum load is controlled.
- The by-pass valve will be used to control airflow to the boiler modules.
- The gas turbine can be controlled as an air compressor providing air to the boiler modules and no electrical power.

The following normal control functions will be performed, but not necessarily in the same manner as in the W501:

- Start-up
- Overspeed
- Surge
- Overtemperature
- Minimum load
- Minimum and maximum fuel limits
- Automatic synchronization
- Safety protection for vibration, bearing temperatures, generator temperatures, lubrication oil system, and others as required
- Fuel transfer, oil-gas, if required
- Normal shutdown

The gas turbine will be started up and accelerated to 3600 rpm by using the starting motor and by-pass combustor. Fuel is scheduled to give the proper acceleration rate to maintain the turbine within the temperature limits and the compressor away from the surge line. The start-up will be smooth and with little danger of surge or overtemperature because of the smaller combustors, which permit flame stability at a lower fuel flow. This will cause less temperature rise at low airflow. When the unit has reached 3600 rpm, the combustor will be automatically adjusted to maintain a minimum load of approximately 4 MW. From this point on in the loading schedule, the gas-turbine fuel will provide only enough heat to prevent the gas-turbine load from dropping below 4 MW. As the heat input to the boiler module increases, the gas-turbine fuel will decrease and then stop until it is again needed for unit shutdown, at which time it will be restarted.

Surge protection will be provided by operating the fuel valves (if in service), positioning the inlet guide vanes, and opening the compressor bleed valves.

Control of overtemperature will be accomplished by reducing fuel flow to the combustor if it occurs during start-up, and by running back the fuel to the boiler modules if it occurs on load control.

Since total airflow control will be required by the boiler modules and accomplished by positioning the compressor inlet guide vanes, several of the control variables normally associated with gas turbines will be eliminated.

The by-pass airflow valve will be an addition to normal gas-turbine controls and will be used to control airflow to the boiler modules during start-up by permitting air to by-pass the boiler module.

This by-pass air will flow through the combustor and mix with the airflow from the boiler modules.

One other major difference will be that turbine inlet temperatures can be measured and limited directly rather than estimated by using blade path temperature, exhaust temperature, and combustor shell pressure.

A variable pressure drop will exist between the compressor discharge and the turbine inlet, and both values will be used in control and protection of the gas turbine. This pressure drop varies because of the boiler module operation.

The gas turbine can act as an air compressor rather than an electrical generator for boiler module shutdown if no electrical power is required. This will be accomplished by controlling the fuel to the combustor to adjust the turbine speed as needed to maintain the desired pressure. The inlet guide vanes will be used in conjunction with variable speed to provide the desired airflow and to keep the gas turbine in the proper speed range.

Overspeed protection for the gas turbine will be provided by interlocking the gas-turbine overspeed safety system with the boiler module fuel valves, operating the compressor bleed valves, and positioning the inlet guide vanes. In order to prevent excessive overspeeding of the gas turbine in the event of the loss of its electrical load, pressure relief valves are necessary in the gas-turbine piping loop. The gas turbine is designed to withstand 10% overspeed. Therefore, the pressure relief valves should be sized so as to give zero acceleration of the turbine rotor at 10% overspeed with the relief valves open and a gas temperature of 1600°F. The overspeed trip valve can be located in the air line between the compressor and the fluidized bed boiler module as shown in Figure 105, or it can be located in the hot products line between the fluidized bed boiler and the gas turbine.

For cold side installation, the calculated valve sizes are 35 in. for one valve and 25 in. for two. Since the diameter of the four air pipes between the gas-turbine manifold and the boiler module is 28 in., the use of a 25-in. valve in each of two 28-in. diameter pipes is indicated to be the optimum arrangement.

For hot side installation, the calculated valve sizes are 45 in. for one valve and 32 in. for two valves. Since the diameter of the two hot gas pipes between the boiler module and the gas turbine is 54 in., the use of a single pressure relief valve would be feasible.

From the standpoint of reliability of operation and ability to restore service quickly, the use of pressure relief valves in the cold side piping is definitely preferred.

The control system inputs required in addition to those normally used for W501 control will be turbine inlet pressure and total airflow, by-pass airflow, and combustor fuel flow. Blade path

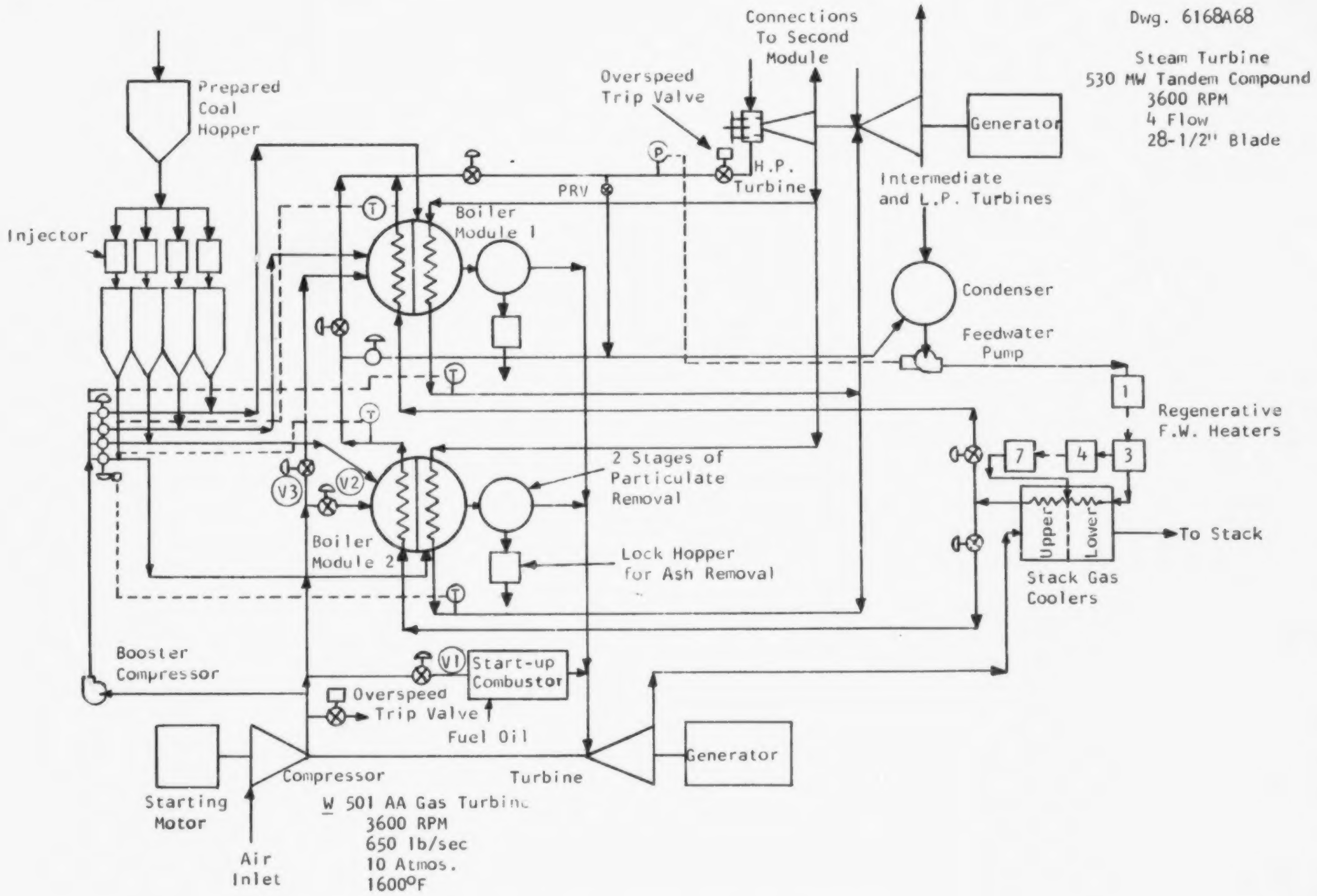
Dwg. 6168A68
Steam Turbine
530 MW Tandem Compound
3600 RPM
4 Flow
28-1/2" Blade
Connections To Second Module
Overspeed Trip Valve
Prepared Coal Hopper
Generator
H.P. Turbine
PRV
Intermediate and L.P. Turbines
Injector
Boiler Module 1
Condenser
Feedwater Pump
1
Regenerative F.W. Heaters
7
4
3
2 Stages of Particulate Removal
V3
V2
Upper
Lower
To Stack
Boiler Module 2
Lock Hopper for Ash Removal
Stack Gas Coolers
Booster Compressor
V1
Start-up Combustor
Overspeed Trip Valve
Fuel Oil
Generator
Starting Motor
Compressor
Turbine
W 501 AA Gas Turbine
3600 RPM
650 lb/sec
10 Atmos.
1600°F
Air Inlet
T
P

Figure 105: Flow Diagram for Plant Start-up

temperature and exhaust temperature measurements are not as critical as usual and will be reduced in the number of measurements.

The outputs required in addition to the normal W501 will be inlet guide vane position demand, bleed valve position demand, boiler module fuel runback, on-line fuel stop and restart signals, and by-pass valve position demand.

The master control will provide start and stop signals, total airflow set point, compressor discharge pressure set point (when operating as an air compressor), and by-pass airflow set point. Mode selection logic can be initiated by the master control as well as by the operator.

5.7.6 Flash Tank

One flash tank will be provided for the four boiler modules and used on only one boiler at a time. The A, C, D, and E valves are associated with the flash tank and will be common to all of the modules. The I, U, N, P, V, W, Y, and B valves will be associated with each boiler module and four sets will be provided, one for each boiler. The boiler by-pass valves are discussed in Section 5.7.8.

The flash tank will be used during clean-up, start-up, and pressure ramping of the once-through boiler modules. When the boiler has reached rated pressure and the P valve is closed, the flash tank is available for use by the next boiler.

The flash tank will function in the normal manner for a sub-critical once-through boiler. The level will be controlled by the D and E valves, the pressure by the A and C valves, and heat recovery by the C and E valves.

The set points for level, pressure, and feedwater temperature will be received from the master controller, since the flash tank is common to the boiler modules and must be coordinated with the plant operation for start-up.

The heat recovery of feedwater temperature control will be interlocked with the actual feedwater temperature so water or steam from the flash tank will not be admitted to the no. 6 or no.7 heaters until the water or steam temperatures are higher than the corresponding heater outlet temperature.

The inputs to the flash tank control system, other than the automation and set points received from the master, will be flash tank level, flash tank pressure, no. 6 feedwater heater outlet temperature, and no.7 feedwater heater outlet temperature.

The output control signals will be valve position demand for valves A, C, D, and E.

5.7.7 Boiler Module

The boilers will be subcritical once-through and controlled in the conventional manner for a plant operating in the turbine-following configuration. That is, a feed-forward system with pressure controlled by the steam-turbine governor valves, load controlled by regulating water and fuel flow, and steam temperature controlled by adjusting the water flow rate to fuel flow rate ratio. Desuperheating sprays between the primary and secondary superheaters will be used for superheat steam temperature control during transient conditions. Reheat steam temperature will be controlled by varying the fuel flow to the reheater bed and by inlet sprays, if the bed is at minimum fuel input. These sprays can only lower the reheat temperature.

Airflow will be held constant to the boiler modules and the individual beds so fuel-air ratio will not be controlled. However, the maximum fuel permitted to any bed will be limited by the airflow to that bed. The bed fuel-air ratio will be determined by measuring the oxygen in the exhaust from each bed.

Since the fuel will be controlled to the four beds of a boiler module individually, the water to fuel ratio to each bed must be maintained. Also, the total water to fuel ratio for the module must be maintained because the fluid flows through the four beds in series.

Since the water and steam flow through the beds will be in series, they cannot be adjusted individually, and the amount of steam produced will be determined by the water flow. The water/fuel ratio must be adjusted by varying the fuel flow. This correction will be done slowly because the pressure will also be effected by changes in water flow or fuel flow.

The desired water or steam outlet temperature of each module will be a function of load and water inlet temperature. These values are known and will be used by the control system to trim the water/fuel ratio of each module by adjusting the fuel.

The upper and lower limits of the fuel flow to each bed will be a function of the bed temperature, 1300°F minimum and 1750°F maximum.

The control system will also be adaptive, in that it will adjust the outlet water or steam temperature, if necessary, to keep the correction of each module near the minimum. Also, the heat absorption will be adjusted between the modules if any of the beds approach the 1300°F minimum or 1750°F maximum. This will be done by adjusting the individual water or steam outlet temperatures so that all of them approach the high or low bed temperature limit at the same time.

The steam temperature control used for transients will be a cascaded control system that maintains a differential temperature by varying the spray flow. The spray flow will be held at approximately 3 to 5% of boiler capacity for normal operations so the temperature can be controlled for either increasing or decreasing loads and/or temperatures.

Water flow to the boiler in other than the first module will be controlled by the B valve; on the first boiler module it will be controlled by the feedpump.

Coal will be fed to the bed by positioning four control valves in parallel.

The stone will be recirculated as needed for regeneration, and added or removed as required to maintain the desired bed depth.

The inputs to the control system for each boiler module will be water flow, water-wall outlet temperature, first superheater outlet temperature, second superheater outlet temperature, reheater outlet temperature, desuperheater inlet and outlet temperature, O_2 content of each bed exhaust gas, temperature of each bed, depth of each bed, coal flow to each bed, igniter fuel flow to each bed, SO_2 of each bed exhaust gas, and particulates leaving the collectors.

The outputs from the control will be four fuel demands, four igniter fuel demands, four bed air valve position demands, reheat valve position demand, B valve position demand, superheater spray valve position demand, and reheat spray valve position demand.

5.7.8 Boiler By-Pass

Each boiler module will have a set of valves for start-up. These will be in addition to the flash tank valves described in Section 5.7.6 and will receive their set points from the master control. The module start-up will be coordinated with the rest of the plant. The valves will perform the following functions:

- I valve. Isolates the boiler from the turbine and also controls steam flow while bringing the boiler onto and off the line
- U valve. A back pressure controller. It will maintain the steam pressure in the second superheater by spilling steam to the condenser. It can also act as a relief valve while the boiler is on-line.
- V valve. A stop valve between the first and second superheaters. It will be positioned to transfer from the flash tank to on-line.

- N valve. An open-closed valve used to admit steam from the flash tank to the second superheater. The valve will be held closed until the steam pressure in the flash tank is sufficient to be admitted to the second superheater.
- P valve. Admits water or steam from the primary superheater to the flash tank. It will be positioned with the V valve to transfer from the flash tank to on-line.
- W valve. Used to control water-wall pressure during start-up.
- Y valve. A large valve in parallel with the W valve and opened when the unit is on-line.
- B valve. Used to control water flow as described in Section 5.7.7.
- H valve. The H valve will be an isolation valve.

The input signals to the control system will be feedwater flow, water-wall pressure, flash tank pressure, second superheater outlet temperature, second superheater output temperature, turbine header pressure, turbine header temperature, water-wall outlet temperature, and feedwater temperature.

The outlet signals will be valve position demands for the W, Y, P, N, V, U and I valves.

5.7.9 Regeneration System

The regeneration system described is assumed to provide a basis for an evaluation of a total system. Alternative operating features and systems are being studied.

Typical areas which require further evaluation are:

- Reducing gas, producer-gas composition, reliability and flexibility
- Temperature control

- Interface with variable pressure in boiler
- Turndown-single unit versus separate unit for each boiler or turbine module
- Stone circulation system operability.

For control purposes, the specific regeneration system being considered will be divided into three subsystems:

- Stone feed system
- H_2S generation system
- $CaSO_4$ reducing system.

Each of these subsystems will be controlled essentially independently of the others, provided the other systems are in operation. The interaction between the systems will be primarily the feed of stone from one subsystem to the next and will be primarily affected by the amount of sulfur contained in the stone. This cannot be measured but can be estimated from empirical information by knowing the amount of coal feed and the sulfur content of the coal. For this reason, the feed-forward system will be used with corrections made to each subsystem to properly balance the regeneration system with itself and with the remainder of the plant. For steady-state operation, load demand will be proportional to coal feed and used as the feed-forward signal. Boiler modules can operate once-through during start-up and for short periods on-line. The feed-forward signal will be corrected for the number of boiler modules shut down or operating once-through and their portion of the load.

5.7.10 Stone Feed System

The stone will be fed to the 20 beds by pneumatic conveyors from a common surge vessel. Make-up stone will be fed to the surge vessel at the rate of approximately 15% of the stone feed rate by means of a double bell feeder. If the level drops below a predetermined value, the make-up rate will be increased. Overrides will be provided to permit raising the level in the surge vessel for start-up. The make-up feed will be stopped if the surge vessel should fill to a predetermined high level.

Each boiler module will have a hold vessel which will be controlled to a maximum level by removing spent stone. This control will be of the on-off type operating between two levels. Overrides will permit the hold tank level to be lowered to a minimum value for start-up.

Stone will be fed to the beds at a fixed rate by utilizing variable time control of the pneumatic feed system. SO_2 will be measured in the exit of each boiler module. If the SO_2 content increases to a preset value, the sampling point will be moved to each bed to determine which bed is producing the SO_2, and the stone feed will be increased to that bed. When the system is again in balance, the SO_2 measurement will be returned to the boiler module exit.

Stone will be removed from each bed by means of a weir which keeps the level constant. Each bed will also contain a valve which can be used to lower the bed depth for start-up.

The master signals to the stone feed system will be stone feed automation commands, number of beds in service, and number of beds operating once-through.

The measured inputs will be surge vessel level, four hold vessel levels, four SO_2 percentages leaving boiler or bed, and 20 bed levels.

The outputs will be control signals to surge vessel make-up, four spent stone removal valves, 20 stone feed to bed valves, 20 stone from bed valves, and SO_2 sampling point selection.

5.7.11 H_2S Generator Control

The H_2S generator control system will be a feed-forward system that controls the CO_2 and H_2O flows as a function of load demand with corrections for percent H_2S in the exit gas and humidity of the CO_2 to the H_2S generator vessel. Pressure in the H_2S generator vessel will be controlled constant, and a surge control system will be provided to protect the compressor during start-up.

The load demand signal will be corrected for the number of modules shut down or operating once-through. This corrected signal demands a position of the expander throttle valve after being trimmed by a signal proportional to the deviation of compressor suction pressure from the desired set point. There will be two expander throttle valves: one to regulate air from the gas-turbine compressor and one to regulate gas from the $CaSO_4$ reducer vessel. The air throttle valve will be used only during start-up. The suction pressure deviation trim signal will be used during start-up.

The CO_2 flow demand will be corrected to produce the desired H_2S in the discharge gas from the H_2S generator vessel. CO_2 production will be greater than that required by the process and controlled by positioning the stack gas control valve and the stripper control valve. Actual CO_2 flow to the H_2S generator vessel will be controlled by positioning the CO_2 bleed valve. This configuration will permit an accurate control of CO_2 flow and permit the system to handle transient conditions. The percent H_2S trim is not used during start-up.

The H_2O to CO_2 ratio will be controlled by maintaining a desired condenser discharge temperature. The condenser will produce saturated gas. The remainder of the H_2O required to produce the desired ratio will be provided by adding steam to the CO_2 at the compressor discharge. This permits accurate control during load transients and steady state operation. It also keeps the power required to operate the compressor to a minimum.

The pressure in the H_2S generator vessel will be maintained at the desired value by modulating the discharge valve.

Three bleed valves will be required to protect the compressor from surge during all operating conditions. Bleed valves no. 1 and no. 3 will be used during start-up, but cannot be used on-line because air must be kept out of the system. Bleed valve no. 2 will provide the recirculation of CO_2 to maintain minimum flow without admitting any air to the system. Minimum flow for safe operation will be determined from compressor pressure ratio and speed.

The master signals to the H_2S generator system will be automation commands, load demand, beds in service, and beds operating once-through.

The measured inputs will be exit gas percentage CO_2, H_2O and H_2S, H_2S generator vessel pressure, CO_2 flow, compressor pressure ratio, compressor speed, and compressor inlet pressure.

The outputs will be control signals to the expander air throttle valve; expander gas throttle valve; stripper flow valve; condenser cooling water valve; steam valve; H_2S generator vessel discharge valve; compressor bleed valves no. 1, no. 2, and no. 3; stack gas control valve; CO_2 bleed valve; CO_2 isolation valve; vent valve; and two stone isolation valves.

In order to maintain fluidization in the vessels during reduced load, multiple regeneration modules may be required rather than the single module concept discussed here.

5.7.12 $CaSO_4$ Reducer Vessel

The $CaSO_4$ reducer control system will be a feed-forward system that controls the gas from the gas producer and steam to the $CaSO_4$ reducer vessel as a function of load demand, with corrections for percentage combustibles in the exit gas and exit gas temperature. Pressure in the $CaSO_4$ reducer vessel will be controlled constant.

The corrected load demand signal discussed in the H_2S generator system will be used in this system. It positions the air and coal feed after being corrected to maintain the desired percent combustibles in the exit gas. The load demand signal to the steam valve will be corrected for exit gas temperature.

The pressure in the $CaSO_4$ reducer vessel will be maintained constant by positioning the expander by-pass valve.

The master signals to the $CaSO_4$ reducer system will be automation commands, load demand, beds in service, and beds operating once-through.

The measured inputs will be exit gas percentage combustibles, exit gas temperature, reducer vessel pressure, gas producer steam flow, gas producer airflow, and gas producer coal feed.

The outputs will be control signals to gas producer steam flow, gas producer airflow, gas producer coal feed, and expander by-pass valve.

5.8 CONCLUSIONS AND COMMENTS

5.8.1 Conclusions

- The plant should be capable of control for base load or swing operation.
- By changing the number of boiler modules and gas turbines in operation, the plant can operate over a load range or from 12-1/2% to 100% plant capacity. This is comparable with conventional power plants.
- Load changes of 5% per minute over the load range of the equipment in service can be accomplished, that is from 50% load to 100% load if all four modules are operating. This rate of change is comparable with other types of large power plants.

5.8.2 Comments

- New procedures for start-up of the once-through boilers can be developed instead of using the conventional systems. This is possible because of the multiboiler configuration and the ability to by-pass steam or water to the condenser. Also, since the minimum water flow per bed is approximately 40%, the start-up valves can be simplified.
- The control system must be of the feed-forward type with corrections for pressures, temperatures, and flows. Since the fuel to each portion of the boiler can be controlled independently (that is preevaporator, primary superheater, the secondary superheater), the temperature and pressure controls should be capable of preventing excursions normally associated with load change transients.
- At least two different loading procedures are required. One for known final conditions and one for ramp-type increases without known endpoints. This is necessary to prevent transients on individual modules.
- More automation will be required than for the conventional plant because of starting and stopping modules to change load.
- The chemical recovery control and operation must be coordinated with the total plant operation. Once-through operation of the stone permits the plant to operate at low loads without the chemical recovery plant.
- Since the fuel can be controlled to each portion of the boiler independently, the temperatures within the boiler can be controlled. This feature will permit better steam temperature control.

- The short resident time of the water in the boiler minimizes the pressure and temperature transients normally associated with once-through boilers.
- Airflow measurement to each bed will be difficult. The damper position of the various beds can provide an approximate measurement. Gas analysis can be used to determine excess air so airflow measurement may not be critical.
- The sampling systems for the O_2 and SO_2 measuring equipment will require high maintenance because of the dirty gas at the point of measurement.
- A gas turbine hot gas by-pass will be required to provide for a loss of load on the gas-turbine generator. This would permit the boilers to operate momentarily with the gas-turbine generator out of service.
- The condenser and desuperheater design and control should be investigated to assure that steam can be by-passed to the condenser for boiler module start-up and shutdown.
- The stone flow control and bed depth measurements will be a problem. The stone feed problem is also effected by the variable pressure operation of the beds caused by the changes to gas-turbine inlet pressure as a function of load.
- A combustor may be required to start the expander used by the chemical recovery plant.
- Superheater and reheater spray valves will be required to increase the operating range at constant steam temperature and to provide better temperature control during transients.
- Surge control will be required for gas-turbine compressors. This system will have to be more complex than the open-loop system normally used on gas turbines. It should utilize pressure ratio, inlet volume, and speed for control.

6. ALTERNATIVE PRESSURIZED FLUID BED COMBUSTION CONCEPTS

Fluidized bed combustion technology can be applied with many different configurations and can be utilized in many different steam and power cycles as previously indicated. Two alternative pressurized fluid bed combustion concepts were studied -- an adiabatic combustor combined cycle plant which represents a modification of the base power cycle and a recirculating bed boiler which represents an alternative fluidized bed boiler concept. Conceptual designs, performance, and economics are projected and compared with the pressurized fluid bed boiler combined cycle plant.

6.1 FLUIDIZED BED ADIABATIC COMBUSTOR COMBINED CYCLE POWER PLANT

6.1.1 Adiabatic Combustor Concept

A pressurized fluidized bed boiler combined cycle power plant has been designed using state-of-the-art power generation equipment.[1] Performance, costs, and pollution abatement were projected for the system. The results show the concept has the potential to meet SO_2, NO_x, and particulate emission standards and may reduce energy costs 10% below a conventional plant with stack gas cleaning. This is achieved by effectively combining the combustion, heat transfer, and pollution control processes. The fluidized bed boiler operates at excess air values from 10 to 100% with up to 70% of the heat released in burning the fuel with air transferred to the water/steam in the tubes surrounding and submerged in the bed.

Increasing the design point excess air with constant bed temperature will decrease the total heat transfer surface in the fluid bed until no boiler tube surface will be required at an excess air of approximately 300%. In this case, the power system is a combined cycle plant with the gas to the turbine expanders supplied from a coal-fired, adiabatic combustor.[2]

Combined cycle plants of this type, which burn natural gas and/or heavy distillates, are now being marketed to electric utilities for intermediate and base load applications. One embodiment of such a plant is the Westinghouse Power at Combined Efficiencies (PACE) plant. In the PACE plant, the exhaust from the gas turbine is reheated to a temperature of about 1200°F ahead of the heat recovery boiler which has steam conditions of 1250 psia/950°F.

A modified PACE configuration, shown in Figure 106, was selected for evaluation of the adiabatic combustor system. In this configuration, the heat recovery boiler is unfired and desulfurization is achieved by feeding limestone or dolomite to the adiabatic combustor with once-through sorbent utilization.

6.1.2 Cycle Performance

Process conditions for four alternative configurations are summarized in Table 44. Case IV presents the generating and performance data for the unfired heat recovery boiler case selected for evaluation.

As in the high-pressure fluidized bed boiler, the desulfurization process limits the maximum temperature in the fluidized bed to about 1750°F, which means that the maximum gas-turbine design temperature is about 1700°F. With a 1700°F gas-turbine inlet temperature, the temperature of the exhaust gas to the heat recovery boiler is about 850°F. This precludes the use of the standard PACE plant steam conditions. The steam conditions selected for the adiabatic combustor system are approximately 700 psia/800°F. Also, since the gas temperature to the boiler will decrease with the gas-turbine load, it will be necessary to use a sliding pressure boiler in which the steam conditions will follow the exhaust gas temperature at part load. The estimated heat rate at the design point for this modified PACE plant is 9096 Btu (HHV)/kWh.

The standard PACE plant consists of two W501 gas turbines with a modularized heat recovery boiler for each gas turbine. With

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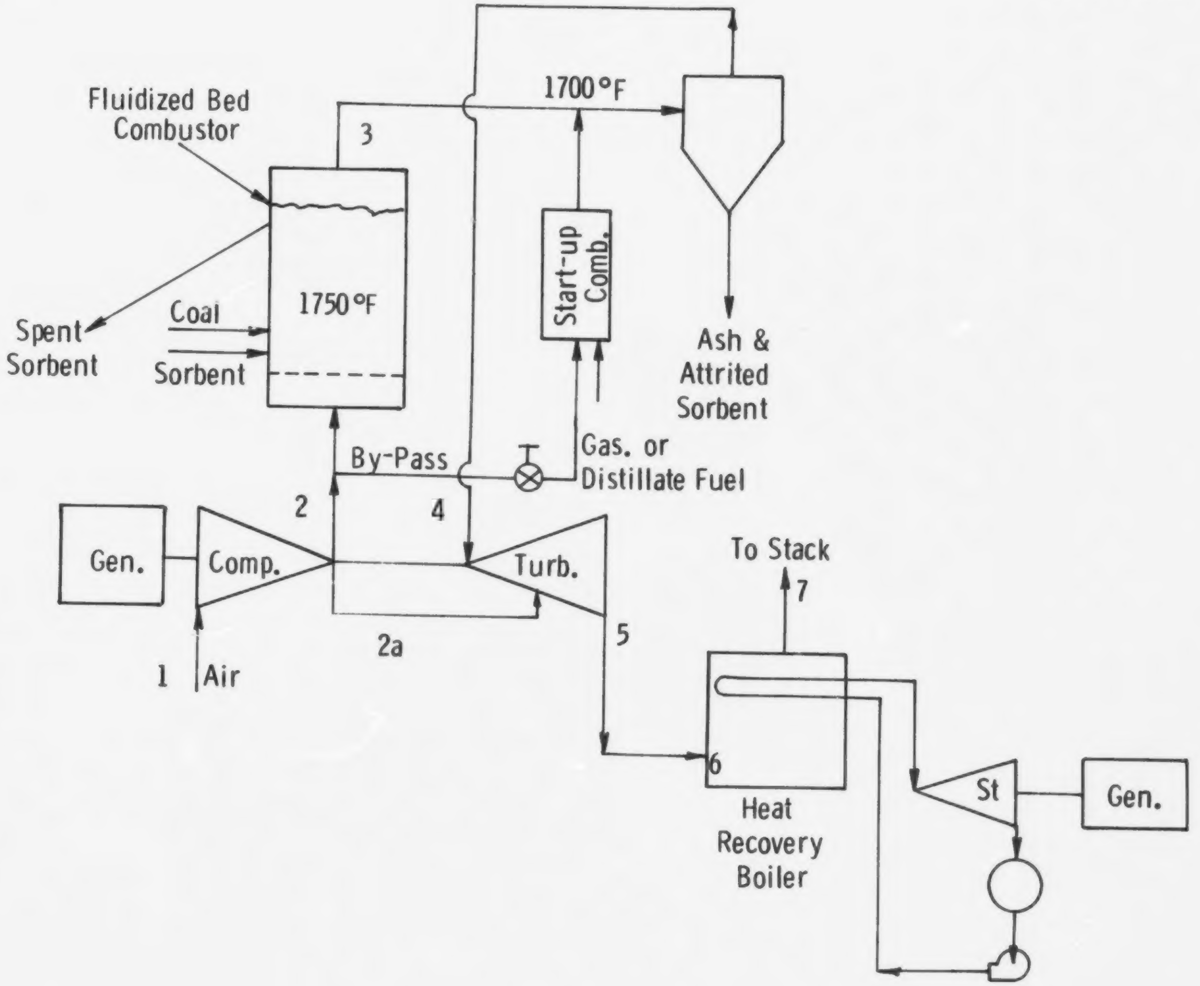

Fluidized Bed Combustor
3
1700 °F
1750 °F
Spent Sorbent
Coal
Sorbent
Start-up Comb.
Ash & Attrited Sorbent
By-Pass
Gas. or Distillate Fuel
2
4
Gen.
Comp.
Turb.
To Stack
7
5
2a
1
Air
6
Heat Recovery Boiler
St
Gen.

Figure 106: Coal-Fired PACE Plant

TABLE 44

PROCESS CONDITIONS AND PERFORMANCE

	Case			
	I	II	III	IV
t_1 - °F	59	59	59	59
W_{a1} - lb/sec	1535	1535	1535	1535
p_1 - psia	14.65	14.65	14.65	14.65
p_2 - psia	146.5	146.5	146.5	146.5
t_2 - °F	614	614	614	614
W_{a2} - lb/sec	1450.0	1429.5	1450.0	1429.5
t_{2a} - °F	450	450	450	450
P_{2a} - psia	146.5	146.5	146.5	146.5
W_{2a} - lb/sec	81.9	102.3	81.9	102.3
p_3 - psia	135.5	135.5	135.5	135.5
t_3 - °F	1600	1700	1600	1700
W_{p3} - lb/sec	1485	1468.5	1485	1468.5
p_5 - psia	15.2	15.2	15.06	15.06
t_5 - °F	798	860	795	857
W_{p5} - lb/sec	1567	1570.6	1567	1570.6
t_6 - °F	1200	1200	795	857
p_6 - psia	15.06	15.06	15.06	15.06
W_{p6} - lb/sec	1583	1584.3	1567	1570.6
t_7 - °F	359	359	411	408
P_s - MW	136.5	136.8	49.6	60.5
P_{GT}	126.9	141.2	128.1	142.4
P_G	263.4	277.9	177.6	202.9
P_N	259.0	273.6	175.4	200.4
HR(HHV)	9281	9003	9440	9095
W_{f1} - lb/sec	35	39	35	39
W_{f2} - lb/sec	16	13.7	0	0

the gas turbine and steam conditions described above, this modified PACE plant would have net power output of 200.4 MW. In order to compare this concept with the high-pressure fluidized bed boiler with a capacity of 635 MW, a plant consisting of three coal-fired PACE systems was assumed with a total capacity of 601.2 MW.

A comparison of heat rates on the same basis for the various coal burning power plants is as follows:

Type of Plant	Heat Rate - Btu (HHV)/kWh
Conventional Steam	9186
High-Pressure Fluidized Bed Boiler	8990
Coal-Fired Modified PACE Plant	9096

6.1.3 Plant Turndown

The turndown capabilities of the coal-fired modified PACE plant should be very wide. Figure 107 shows that a reduction of about 40% is possible by variation of bed temperature from 1750 to 1400°F (corresponds to turbine inlet temperatures of about 1700 and 1350°F). Modulation of compressor air provides another 20% turndown. Beyond this point, compressor air can by-pass the combustor (see Figure 106) operating at constant bed temperature to reduce the gas-turbine inlet temperature to the idle level of about 950°F. It would probably not be practical to operate the boiler all the way to this point. Therefore, the boiler could be shut down and the plant power reduced to near zero. A by-pass around the boiler might be necessary. In order to carry out the by-pass technique described above, the design superficial velocity would have to be high enough to permit about 30% reduction in flow at a bed temperature of 1400°F.

A general discussion of plant turndown for pressurized fluid bed combustion boiler power plants is presented in Section 5.3 of this report.

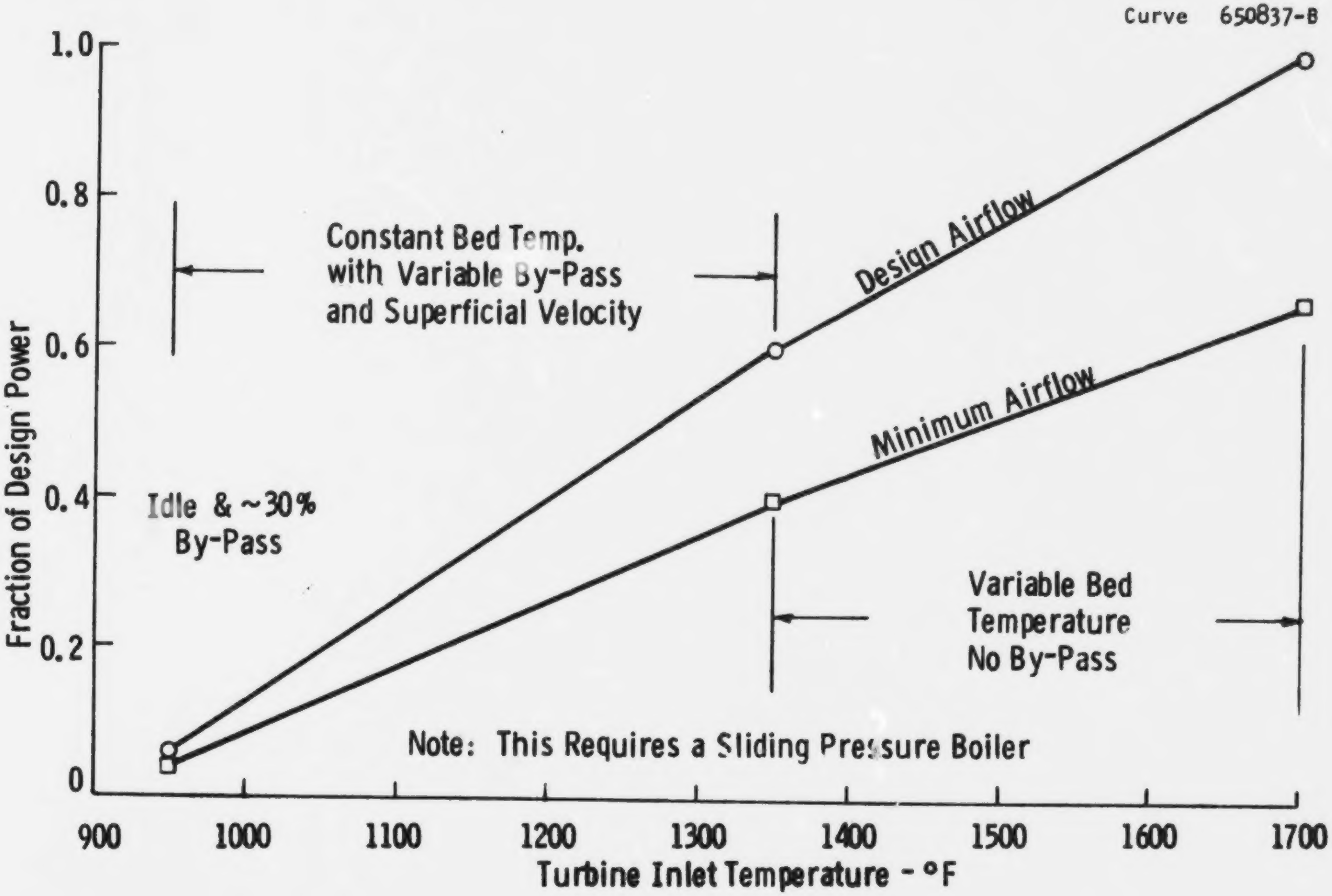


Figure 107: Turndown Capabilities of Coal-Fired PACE Plant with Unfired Heat Recovery Boiler

6.1.4 Environmental Considerations

The environmental impact of an adiabatic fluidized bed combustor power plant is projected in Table 45 . Sulfur dioxide emission standards can be met based on the available pressurized fluidized bed boiler data and the kinetic data at the adiabatic combustor operating conditions (see Section 3.10). Data are required to determine if the nitrogen oxide emission standard can be met. Nitrogen oxide limits can be achieved in a pressurized fluid bed boiler. However, the high excess air for an adiabatic combustor may affect the NO formation. For example, if NO is reduced by reaction with CO, the high excess air may reduce the available CO and increase the NO emission. Particulate emission standards can be achieved in an adiabatic combustor system based on fluid bed combustion emission and particulate removal equipment data.[1] Data on particulate control for an adiabatic combustor system are required to confirm this projection. The gas flow for an adiabatic combustor system is ∿ 40 lb gas/lb fuel compared with ∿ 12 lb gas/lb fuel for a pressurized fluid bed boiler system. This requires a more extensive particulate removal system for the adiabatic combustor system than for a pressurized fluid bed boiler system.

Heat rejection with an adiabatic combustor system will be reduced ∿ 35% below a conventional coal-fired plant. This results from the increased fraction of gas-turbine power (see Table 44). Solids from the system are projected from the pressurized boiler and TG data (see Section 3). Low-grade fuels and solid wastes can be effectively utilized in a fluidized bed adiabatic combustor system.

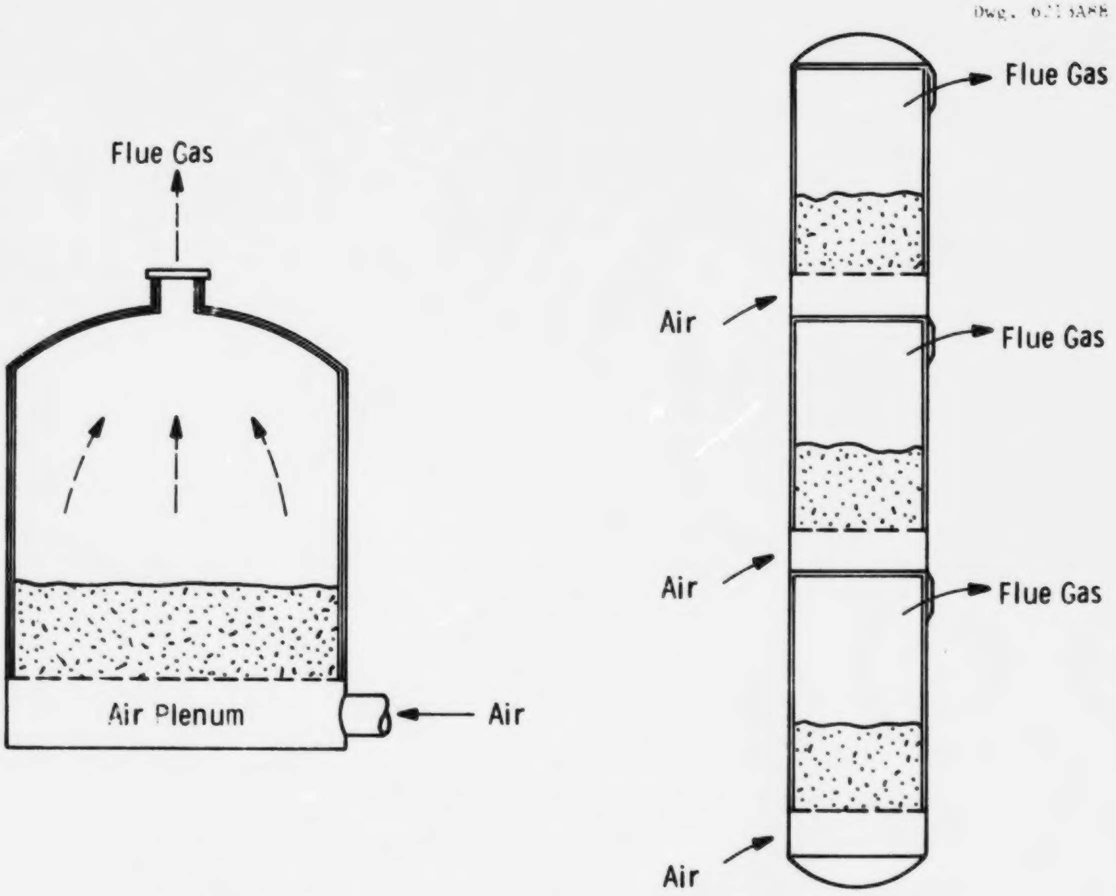
6.1.5 Fuel Processing Equipment

A preliminary evaluation was made of the adiabatic combustor fuel processing system. The design basis is summarized in Table 46. Two design concepts -- a single fluid bed module and a stacked fluid bed module -- were considered as illustrated in Figure 108. A summary of the respective design features is presented in Table 47. The partic-

TABLE 45

ENVIRONMENTAL IMPACT

Item	Impact
Air Emissions	
SO_2, lb/10^6 Btu	< 1.2
NO_x, lb NO_2/10^6 Btu	to be determined
Particulate, lb/10^6 Btu	< 0.1
Heat Rejection to Cooling Water	∿ 35% less than conventional coal plant
Solids	dry, granular $CaSO_4$
Resources	multifuel capability, e.g., low-grade coals, oils, solid wastes


Dwg. 6213A88
Flue Gas
Air Plenum
Air
Flue Gas
Air
Flue Gas
Air
Flue Gas
Air

Single Bed Design

Stacked Bed Design

Figure 108: Adiabatic Combustor Designs

TABLE 46
DESIGN BASIS

- 30 mole % Ca utilization of the dolomite
- 90% removal of SO_2 from burning Pittsburgh #8 coal of 4.3% S

  Case IV 1070 ppm → 107 ppm
- 3 hr average residence time of dolomite
- Refractory insulation

  1700°F normal 2000°F design

  ΔP across wall is minimum (say 0.5 psi)

  $\left\{\begin{array}{l}\text{3-1/2" insulating "A.P. Green VSL50" plus}\\ \text{2-1/2" "A.P. Green Lo Abrade"}\end{array}\right.$

• Dolomite feed rate	179,000 lb/hr (as $CaCO_3$ $MgCO_3$)
Coal feed rate	140,000 lb/hr

TABLE 47

ADIABATIC COMBUSTOR DESIGNS

	Single Bed Design		Stacked Bed Design	
	Design I	Design II	Design I	Design II
Number of Modules	4	2	4	2
Number of Beds per Module	1	1	3	6
Module Diameter, ft	21	30	12	12
Module Height, ft	16	16	50	100
Bed Depth, ft	6.5	6.5	6.6	6.6
Bed Area, ft^2	347	695	113	113
Fluidizing Velocity, ft/sec	6	6	6.2	6.2
Equipment Cost, $/kw	2.5	2.1	2.0	2.1

ulate removal system design is specified in Table 48. A preliminary fuel processing equipment layout is presented in Figure 109 for a single bed module concept.

6.1.6 Power Plant Cost

The fluidized bed boiler power plant cost breakdown presented by Westinghouse was reviewed and compared with an adiabatic combustor power plant. Table 49 presents a comparison of the power plant equipment for the adiabatic combustor plant with the pressurized boiler. Equipment costs for a 600 MW adiabatic combustor system have been projected for two cases:

- Pressurized boiler basis. Assumes six gas turbines, one adiabatic combustor per turbine, and one, 180 MW, steam turbine. Costs are based on the pressurized fluid bed boiler plant costs.
- Modified PACE plant. Assumes three PACE plants but also assumes same style plant as in the pressurized boiler plant, e.g., buildings, stack, etc. Does not account for potential cost increases in undistributed costs due to increased engineering and field supervision as the result of the coal processing system.

An equipment cost summary is presented in Table 50. Equipment costs for the fluidized bed combustion boiler and a conventional plant with stack gas cleaning are also presented. Component costs are presented in Tables 51 and 52 for the adiabatic combustor system. Total power plant capital costs for each system are presented in Table 53.

6.1.7 Conclusions

- Equipment costs for an adiabatic combustor power plant will be less than 4% below those of a conventional plant with stack gas cleaning using the same plant design basis.
- Total capital cost will be 5 to 15% less than that of a conventional plant with stack gas cleaning using the same plant design basis and construction time from 2-1/2 to 3 years.

TABLE 48

PARTICULATE REMOVAL SYSTEM (200 MW)[a]

	1st Stage	2nd Stage
Actual Gas Flow, ACFM/module	124,500	124,500
Collector Selection	size 635 Model 810 Duclone	model 18,000 Tornado Cyclone
Number of Collectors per Module	5	4
Pressure Drop, in. W.G.	20.9	31
Collection Efficiency, % by weight	85.1	97

[a] This table is for the four-module design. For the two-module design, double the number of cyclones per module.

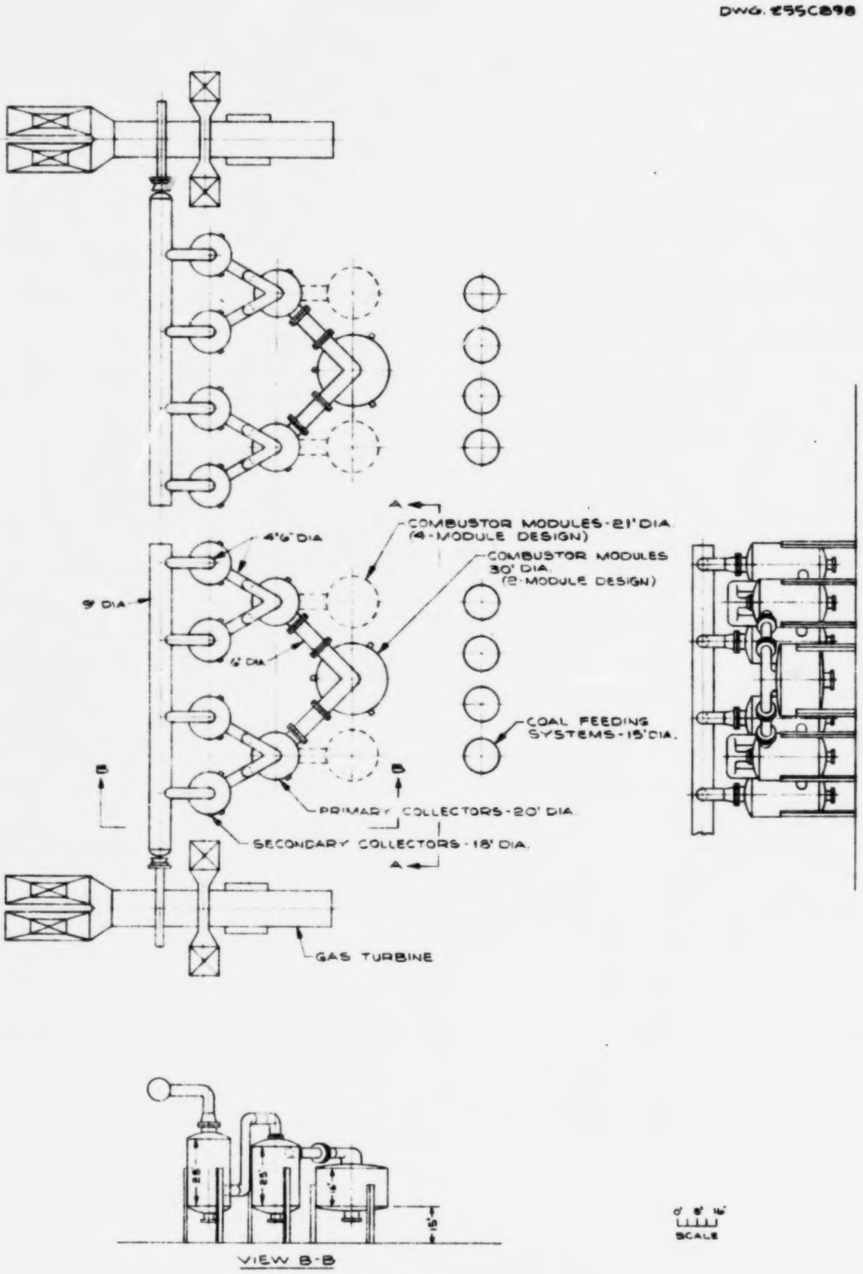

DWG. 255C898
COMBUSTOR MODULES - 21' DIA. (4-MODULE DESIGN)
COMBUSTOR MODULES 30' DIA. (2-MODULE DESIGN)
4'6" DIA
9' DIA
6' DIA
COAL FEEDING SYSTEMS - 15' DIA.
PRIMARY COLLECTORS - 20' DIA.
SECONDARY COLLECTORS - 18' DIA.
GAS TURBINE
A
A
B
B
VIEW A-A
28'
25'
16'
15'
VIEW B-B
0' 8' 16'
SCALE

Figure 109: 200 MW Fluidized Bed Adiabatic Combustor Combined Cycle Plant - Gas Piping from Particulate Collectors

TABLE 49

POWER PLANT EQUIPMENT FOR FLUID BED ADIABATIC COMBUSTOR 600 MW

Structures and Improvements	• structural steel for adiabatic combustors could be reduced - ship with structural steel attached
	• lighting and painting in boiler module area reduced
	• turbine room and heater bay: turbine room may be more expensive if 3 steam turbines used - optimum arrangement may be two modules with two steam turbines to have best steam piping arrangement; heater bay essentially eliminated, although there are 3 deaerators
	• intake and discharge structures cost will be lower since water requirements are $\sim$ 65% less
Turbine Plant Equipment	• circulating water system will be proportionately higher since there are 3 subsystems, one with each PACE module
	• condensing system requires less area and concept does not have feedwater heaters which reduces cost and condensate piping system cost
	• steam turbine: could use one unit or one unit per module; with only one turbine must interface with 6 heat recovery boilers -- tradeoffs not clear
	• feedwater pump, extraction steam piping and steam piping costs are reduced. No hot and cold reheat, piping + lower steam conditions
	• water treatment system less complex due to lower steam conditions
Electric Plant Equipment	• switchgear will be higher since there are more generators
Station Equipment Transmission	• generator voltage

TABLE 50

COAL-FIRED POWER PLANT EQUIPMENT COST COMPARISON

Fluid Bed Adiabatic Combustor

	Conventional Plant[a] with Stack Gas Cleaning 635 MW	Pressurized Fluid Bed Boiler[a] Once-through Dolomite 635 MW	Pressurized Boiler Basis[b] Once-through Dolomite 600 MW	Modified PACE Plant (inc. bldgs. stack, etc.) Once-through Dolomite 600 MW
Land and Land Rights	1.13	0.95	0.95	0.95
Structures and Improvements	27.28	18.52	17.30	9.63[d]
Boiler Plant Equipment or Fuel and Combustion Equipment	82.23	49.68	76.70	76.70
Gas-Turbine Generator Equipment		14.80	46.10	
Steam-Turbine Generator Equipment	55.76	44.14	37.49	
Electrical Equipment	14.99	13.47	13.47	
Miscellaneous Plant Equipment	4.21	3.56	3.56	123.60[f]
Undistributed Costs	30.01	28.35	28.35	
Other Plant Costs	2.91	2.91	2.91	
Total Equipment Cost[c]	218.52	176.38	226.83	210.88[e]

[a] Costs are based on cost data presented in "Evaluation of Fluidized Bed Combustion Process," Westinghouse report to EPA, November 1971. The following cost reductions were made in the fluid bed boiler system: land and land rights cost reduced for once-through system; emergency generator of 22 MW originally projected replaced with 2500 KW generator; four secondary collectors replaced with two larger secondary collectors.

[b] Costs are based on scaling pressurized fluid bed boiler system costs. See Table 47. Concept assumes 6 gas turbines and one steam turbine for a 600 MW plant.

[c] Cost base is January 1970.

[d] Cost reduction results from omission of steam-turbine building (since there are 3, 60 MW units building is not necessary).

[e] See Table 48.

[f] Cost will be higher than indicated due to increased undistributed costs with coal processing system.

TABLE 51

FLUIDIZED BED ADIABATIC COMBUSTOR COMBINED CYCLE POWER PLANT EQUIPMENT COST BREAKDOWN
(based on Fluidized Bed Boiler Plant Costs)

	Equipment Costs	Installation	Installed Costs	
			%	$/kw
Structures and Improvements				
Yard work	432,000	708,000	1,140,000	1.90
Combustor modules[a]	214,000	386,000	600,000	1.00
Turbine room[b]	1,380,000	1,800,000	3,180,000	5.30
Control building	380,000	395,000	775,000	1.29
Services building	1,070,000	1,450,000	2,520,000	4.20
Administration building	385,000	360,000	745,000	1.24
Intake and discharge structures[c]	289,000	521,000	810,000	1.35
Heat recovery boiler[d]	190,000	425,000	615,000	1.02
Subtotal			10,385,000	17.30
Fuel Processing Subsystems[e]				
Adiabatic combustor, two-module design[f]	1,260,000	900,000	2,160,000	3.60
Particulate removal system, 1st- and 2nd-stage collection[g]	18,800,000	3,420,000	22,220,000	37.00
Gas piping from 2nd stage Cyclone to gas turbine[g]	2,880,000	2,400,000	5,280,000	8.80
Air piping from gas turbines to combustors[h]	240,000	120,000	360,000	0.60
Coal feeding and make-up[i]			2,280,000	3.80
Sorbent handling[i]				
Coal handling			6,310,000	10.50
Draft flues and ducts[g]			3,250,000	5.40
Stack and foundation[g]			900,000	1.50
Ash and dust handling[i]			750,000	1.20
Control and instrumentation[i]			2,000,000	3.30
Miscellaneous equipment[i]			600,000	1.00
Subtotal			46,110,000	76.70

TABLE 51 (Continued)

	Equipment Costs	Installation	Installed Costs	
			%	$/kw
Gas Turbine Subsystem				
Modified W501 units (6 @ $4,060,000)	24,360,000			
External manifolds (6 @ $200,000)	1,200,000			
Subtotal	25,560,000	2,100,000[j]	27,660,000	46.10
Steam Turbine Subsystem				
Heat recovery boilers[k]	3,600,000[l]	2,000,000[j]	5,600,000	9.34
Turbine generator[m]	6,070,000[n]	700,000[j]	6,770,000	11.30
Condenser and condensate system[c]	2,120,000	790,000	2,910,000	4.85
Circulating water system[c]	1,065,000	1,120,000	2,185,000	3.64
Feedwater system[o]	1,045,000	675,000	1,720,000	2.57
Other steam-turbine equipment[p]	1,230,000	1,125,000	2,355,000	3.92
Instruments and controls[q]	700,000	240,000	940,000	1.57
Subtotal			22,480,000	37.49
Electrical Subsystem				
Switchgear	695,000	135,000	820,000	1.38
Station service equipment[r]	1,726,000	295,000	2,021,000	3.37
Switchboards	459,000	121,000	580,000	0.97
Protective equipment	90,000	150,000	240,000	0.40
Structures and containers	414,000	1,196,000	1,610,000	2.68
Power and control wiring	869,000	1,931,000	2,800,000	4.67
Subtotal			8,081,000	13.47

---

[a] Reduced structural requirement for reactor vessels.

[b] Reduced heater bay volume requirements.

[c] Reduced throughput.

[d] Foundations for heat recovery boilers.

[e] Total cost could increase ∿ $5/kw if use four combustor modules per turbine instead of two modules.

[f] See Table 47.

TABLE 51 (Continued)

[g] Increased throughput.

[h] Increased complexity.

[i] Omitted from high-pressure fluidized bed boiler.

[j] Supplied by John Kelly of Power Generation Services, Lester, Pa.

[k] Unfired heat recovery boiler
Steam conditions: 695 psia/805°F
Unit capacity: 250,000 lb/hr
Unit cost: $600,000
Total units: 6
Gas-turbine inlet temp.: 1700°F
Gas-turbine exhaust temp.: 850°F

[l] Supplied by David Costa, HTD Marketing.

[m] Steam turbine and generator
Steam conditions: 700 psia/800°F/2.5 in. Hg
Capacity: 180 MW
Turbine configuration: tandem compound
two flow 25 in. ends

[n] Supplied by A. G. Taylor, Power Generation Systems.

[o] Feedwater heaters eliminated except for deaerating heater. Capacity and head for feedwater pumps reduced.

[p] Piping complexity reduced due to elimination of feedwater heaters. Water treatment requirements reduced due to low steam conditions.

[q] Higher fraction gas-turbine power. Increase in number of gas turbines.

[r] Reduced capacity of emergency gas-turbine driven generator. Reduced capacity of transformer for start-up turbine.

TABLE 52

FLUIDIZED BED ADIABATIC COMBUSTOR COMBINED CYCLE POWER PLANT EQUIPMENT COST BREAKDOWN (based on Modified PACE Plant)

	Equipment Costs	Installation	Installed Costs	
			%	$/kw
Structures and Improvements				
Combustor modules[a]	214,000	386,000	600,000	1.00
Fuel Processing Subsystems[b]				
Adiabatic combustor, two-module design	1,260,000	900,000	2,160,000	3.60
Particulate removal system, 1st- and 2nd-stage collectors	18,800,000	3,420,000	22,220,000	37.00
Gas piping from 2nd stage Collector to gas turbine	2,880,000	2,400,000	5,280,000	8.80
Air piping from gas turbines to combustors	240,000	120,000	360,000	0.60
Coal feeding and make-up				
Sorbent handling			2,280,000	3.80
Coal handling				
Ash and dust handling			6,310,000	10.50
Control and instrumentation			750,000	1.20
Miscellaneous equipment			2,000,000	3.30
Subtotal			41,960,000	69.80
PACE Subsystems				
Modified PACE units			72,000,000	120,00[c]
External manifolds	1,200,000	240,000	1,440,000	2.40
Emergency generator[b]	625,000	95,000	720,000	1.20
TOTAL			74,160,000	123.60

---

[a] Foundations for adiabatic combustor modules.

[b] 2500 kw capacity.

[c] Assumes cost of PACE invariant with modified conditions, e.g., no firing of heat recovery boiler which results in lower steam conditions and the use of a 60 MW steam turbine.

TABLE 53

POWER PLANT CAPITAL COSTS -- 600 MW PLANT ($/kw)

	Conventional Plant with Stack Gas Cleaning	Pressurized Fluid Bed Boiler	Fluid Bed Adiabatic Combustor	
			Pressurized Boiler Basis	Modified PACE Plant Basis
Equipment Costs[a]	218.52	176.38	226.83	210.88
Normal Contingency (6%)	13.11	10.58	13.61	12.65
Subtotal	231.63	186.96	240.44	223.53
Escalation[b]	52.12	35.06	36.07 (40.57)	33.53
Subtotal	283.75	222.02	226.51 (281.01)	257.06
Interest during construction	47.88	29.14	25.92 (31.61)	24.10
General Items & Engineering	5.50	5.50	5.50	5.50
TOTAL CAPITAL COST	337.13	256.66	307.93[c] (318.12)[d]	286.66[c]

[a] Breakdown presented in Table 3.

[b] Cost base Jan. 1970
Begin construction July 1971
Construction periods
  Conventional - 4-1/2 years
  Fluid bed boiler - 3-1/2 years
  Adiabatic combustor
    Fluid bed boiler basis - 2-1/2 - 3 years
    Modified PACE basis - 2 to 2-1/2 years
Escalation rate 7-1/2%
Interest rate 7-1/2%

[c] 2-1/2 year construction

[d] 3 year construction

- The capital cost of the pressurized fluid bed boiler plant is up to 20% less than that of the adiabatic combustor plant.
- Plant heat rate can be reduced ∿ 1% compared with conventional coal-fired steam plant using state-of-the-art technology.
- Environmental constraints can be met, heat release to cooling water is reduced ∿ 35%.
- The simplicity of the adiabatic combustor system, the potential for a shorter construction time and the potential for near term commercial application may make the concept commercially viable.

6.2 RECIRCULATING BED BOILER DESIGN

A deep recirculating fluidized bed combustion boiler was conceived as an alternative to the proposed pressurized fluid bed combustion boiler concept which is evaluated in Section 2. A recirculating bed boiler may offer advantages over the proposed design but will require more development and is thus considered a second generation concept

The recirculating fluid bed concept is illustrated in Figure 110. Gas is fed to the base of an open draft tube section. The superficial velocity of the gas flowing up the riser may be 10 to 60 ft/sec. The solids are picked up pneumatically in the draft tube. The effective overall density of solids and gases is less in the draft tube section than in the downcomer, which creates a solid circulation pattern upward through the draft tube and downward in the downcomer. Solids and gases from the draft tube pass into a fluid bed which may be of expanded cross-section. Solids from the fluid bed above the draft tube flow into the downcomers and enter the base of the draft tube. Gas is introduced at the base of the downcomer at a rate necessary to permit the downward flow of solids.

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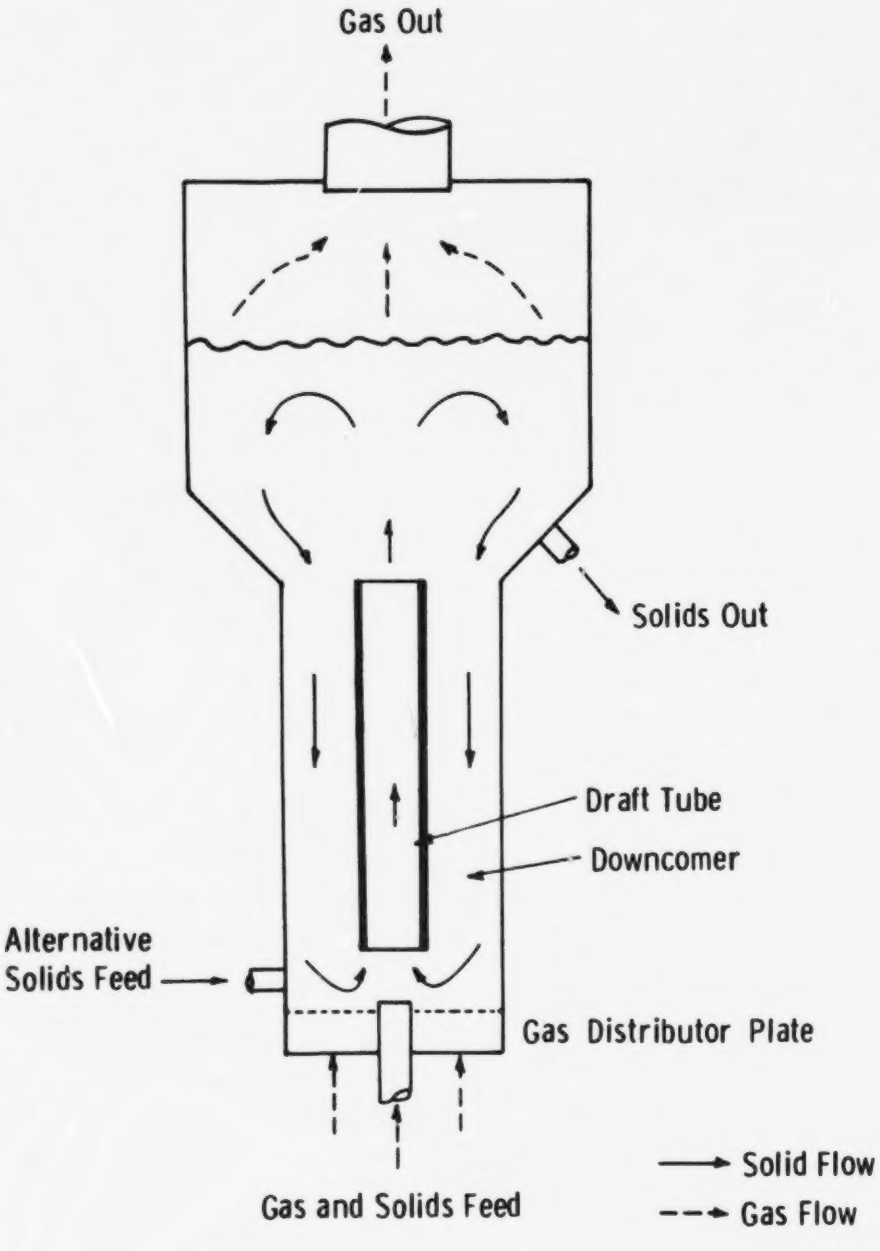
Gas Out
Solids Out
Draft Tube
Downcomer
Alternative Solids Feed
Gas Distributor Plate
Gas and Solids Feed
Solid Flow
Gas Flow

Figure 110—Recirculation Bed Concept

The potential advantages of a recirculating fluid bed boiler concept are:

- Feeding reactants to the bed. Oil and coal have been fed to recirculating beds without distribution and agglomeration problems.
- Boiler turndown and start-up. Locating heat transfer surface in the recirculating solids permits turndown by regulating the solids recirculation rate. Boiler start-up and shutdown are also simplified by so locating heat transfer surface.
- SO_2 removal and NO_x reduction. Limestone or dolomite can be used as the primary bed material in order to remove SO_2 and to catalyze the decomposition of NO. The riser portion of the bed can be operated under reducing conditions: the main bed and recirculating sections, under oxidizing conditions. In this way the formation of NO_x is reduced and the activation of the sorbent is promoted. It is postulated that fluctuation between oxidizing and reducing conditions promotes penetration of the limestone sorbent by the sulfur in the forms of sulfate, sulfite, or sulfide.[3]
- Decreasing the cost of fluidized bed boilers further. A system study of pressurized fluidized bed boiler power plants[1] indicated that they are lower in cost, more efficient in operation, and more effective in pollution abatement than conventional coal- or oil-fired power plants with stack gas scrubbing systems. The deep recirculating beds promise to reduce this cost further because of potentials in decreasing coal feed points, in increasing the heat transfer coefficient in the downcomer, and in simplifying the control and design of the boiler.

Several recirculating fluid bed systems have been developed and operated. The British Gas Council[4,5,6] developed deep recirculating fluidized beds for oil and coal gasification. In the gasifier this was used to smooth out the local high temperatures at the point of oxygen entry and in the hydrogenator to allow for the quick removal of particles wetted by the oil in the vicinity of the oil inlet and thus avoid the formation of agglomerates as well as closely control temperature for process purposes. Experiments were performed in large-scale models up to 5 feet in diameter and operated at pressures and temperatures up to 70 atmospheres and 750°C (1382°F).

The same concept was also used for vertical pneumatic transportation of sticky and bridging powders.[7] A variation of this concept was utilized to promote solid mixing, circulation, heat and mass transfer in the bed[8], and for pretreatment of caking coal[9]. Taskaev and Kozhina[10] utilized a recirculating bed for the low-temperature carbonization of coals. Product gas preheated to a temperature of 600° to 700°C (1112 to 1292°F) was recirculated to entrain the coal particles of sizes up to 1/4 in. through a vertical draft tube 1 in. in diameter and 20 in. in length into a concentric tube 6 in. in diameter and 40 in. in length. The coal or char particles fell through this annular space (a downcomer) in a dense descending bed and were then reentrained by the gas and recirculated until carbonization was complete.

A recirculating fluid bed reactor is utilized in the multistage fluidized bed coal gasification process for production of low-Btu gas for power generation being developed by Westinghouse.[11] A recirculating bed operating near 1600°F is used in this concept to devolatilize the coal and potentially to simultaneously desulfurize the fuel gas.

A preliminary design and evaluation of a pressurized recirculating fluid bed boiler for combined cycle power generation was made. The available literature on recirculating fluid beds was reviewed, cold model studies performed to establish initial operating and design criteria, a preliminary design developed, costs projected, and the concept evaluated.

Based on the above discussions, a deep recirculating fluidized bed appears to be an attractive approach to the design of a boiler. The subsequent sections illustrate the concept of a deep recirculating fluidized bed boiler, describe the cold model studies, and discuss criteria for design and operation.

6.2.1 Deep Recirculating Fluidized Bed Boiler Concept

The concept of a deep recirculating fluidized bed boiler is illustrated in Figure 111. Primary air along with coal or oil fuel is fed to the boiler at the base of an open riser section. This section is bounded by metal or ceramic walls which may contain water-filled tubes for wall cooling and/or steam generation. The superficial velocity of the air and combustion gases flowing up the riser is 10 to 60 ft/sec at the operating temperature and pressure -- 1300°F to 1950°F and 4 to 30 atmospheres. Coal fuel may either be pulverized (less than 200 mesh) or coarsely crushed (perhaps 1/4" x 0). Oil may be introduced to the boiler so as to wet the stream of sorbent after it flows down the downcomer, as it enters the base of the riser and jets upward in the riser. The fuel combines with the primary air in the riser. The heat produced increases the temperature of the combustion products, fuel ash, and sorbent; some heat may be removed through the walls of the riser. The quantity of sorbent entering the base of the riser is controlled by choice of design and operating parameters in order to limit the temperature increase of the gases and solids flowing up the riser.

The solids and gases emerging from the riser pass into a bed of expanded cross-section and decreased superficial gas velocity. Baffles and/or heat transfer surface may be located in the bed section in order to obtain smooth fluidization and recirculation of the sorbent, to improve sulfur removal and NO reduction by the sorbent, and possibly to recover some heat. The gas velocity in the bed is in the range 1 to 15 ft/sec, sufficiently large to fluidize the sorbent but sufficiently small to minimize entrainment. The sorbent may either be coarsely crushed (1/4" x 0) or finely ground (perhaps 100 to 1000μ).

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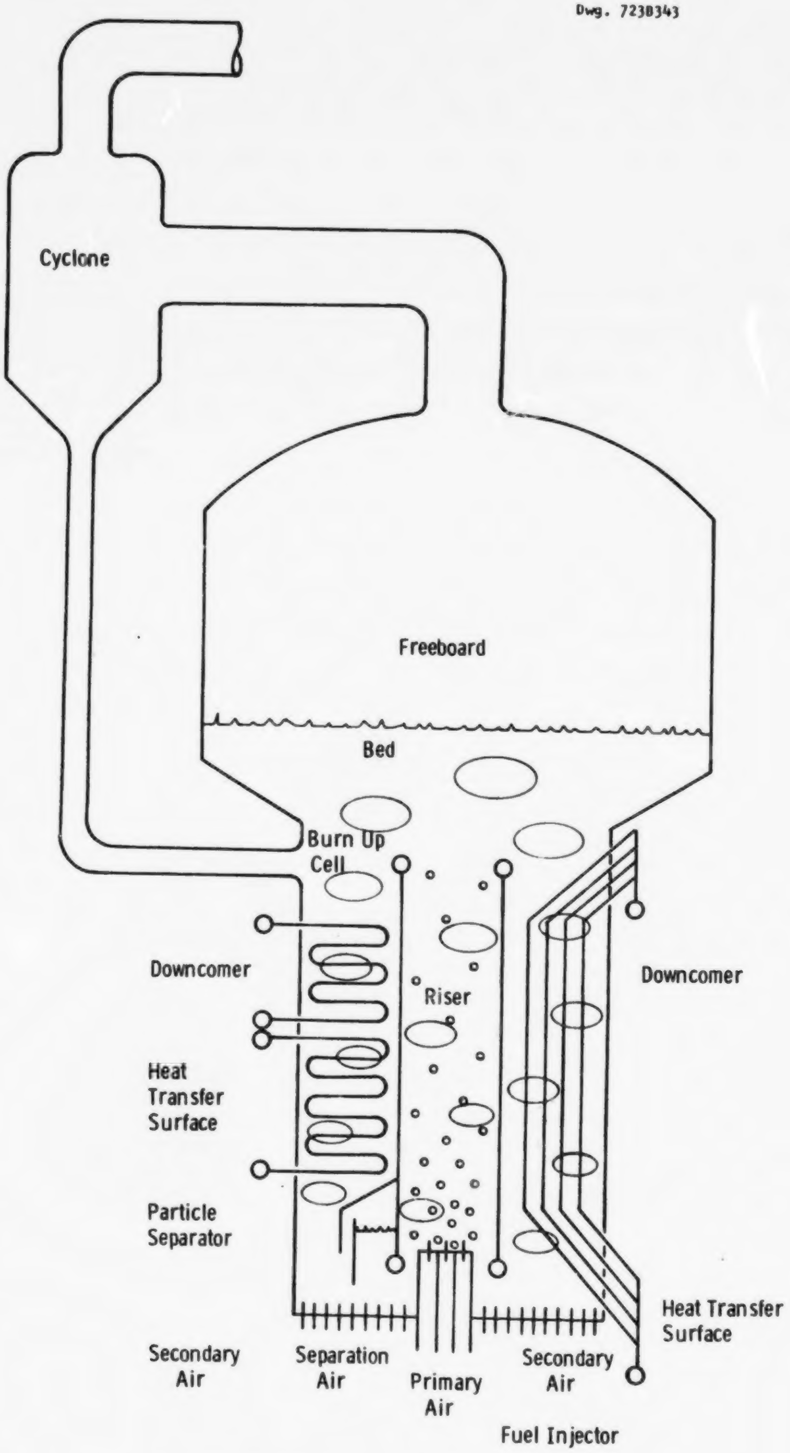
Cyclone
Freeboard
Bed
Burn Up
Cell
Downcomer
Riser
Downcomer
Heat
Transfer
Surface
Particle
Separator
Heat Transfer
Surface
Secondary
Air
Separation
Air
Primary
Air
Secondary
Air
Fuel Injector

Fig. 111–Deep Recirculating Fluidized Bed Boiler

Hot sorbent from the bed flows into the downcomer, passes downward through the heat exchange surface (water/steam filled tubes which may, for example, be platens of horizontal or long vertical tubes), and enters the base of the riser section. The exchange surface extracts the heat energy transferred to the sorbent by the combustion process occurring in the riser and bed. Secondary air is introduced at the base of the riser at a rate necessary to permit the downward free flow of the sorbent and to prevent reducing conditions in the riser due to any residual fuel accompanying the sorbent. Flow of the sorbent in a section of the downcomer can be reduced or halted by reducing or cutting off the flow of secondary air to that section. This capability makes it possible to adjust independently heat removal and heat production in the boiler. Control of the primary air can be used to adjust the recirculation rate of the solids and the degree of oxidation in the riser and above the bed. Combustion after the bed increases the range of gas-turbine operation. Start-up and turndown are thus facilitated.

6.2.2 Cold Model Studies

In order to study the feasibility of using deep recirculating fluidized beds for coal combustion boiler power plants, a two-dimensional cold model was constructed from the transparent acrylic sheets to investigate solid circulation, operation characteristics, and design criteria. The detailed dimensions are shown in Figure 112. Only the bed cross section, 8.5 in. x 1.5 in. is fixed; the other dimensions (A, B, C, D, E, F, and α) in Figure 112 can be adjusted to study the effect of these design variables. Pressure taps are also provided for measuring differential pressure drops between points 1, 2, 3, and 4.

Some experiments were performed in a slightly modified two-dimensional bed as shown in Figure 113. Two separate fluidized beds (or "boots") are created by inserting two aluminum bars 3/4 in. thick underneath the draft tube. The objective of this modification was to study the effect of the draft tube inlet design on the solid circulation rate and the air distribution between the downcomers and the draft tube.

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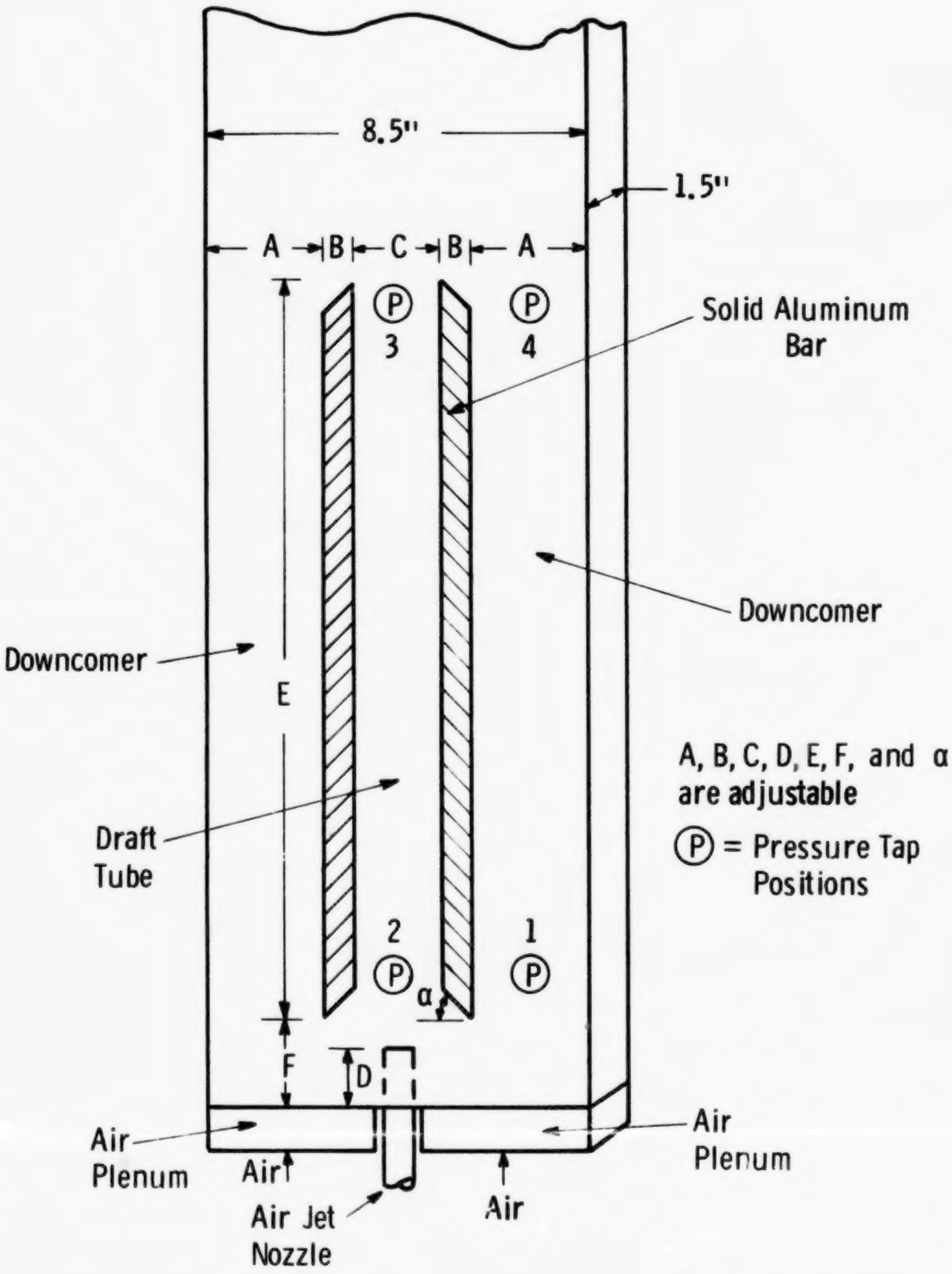
8.5"
1.5"
A
B
C
B
A
P 3
P 4
Solid Aluminum Bar
Downcomer
Downcomer
E
A, B, C, D, E, F, and α are adjustable
Ⓟ = Pressure Tap Positions
Draft Tube
2 P
1 P
α
F
D
Air Plenum
Air Plenum
Air
Air
Air Jet Nozzle

Figure 112—Detailed Schematic of the Two-Dimensional Cold Model

Dwg. 6197A36

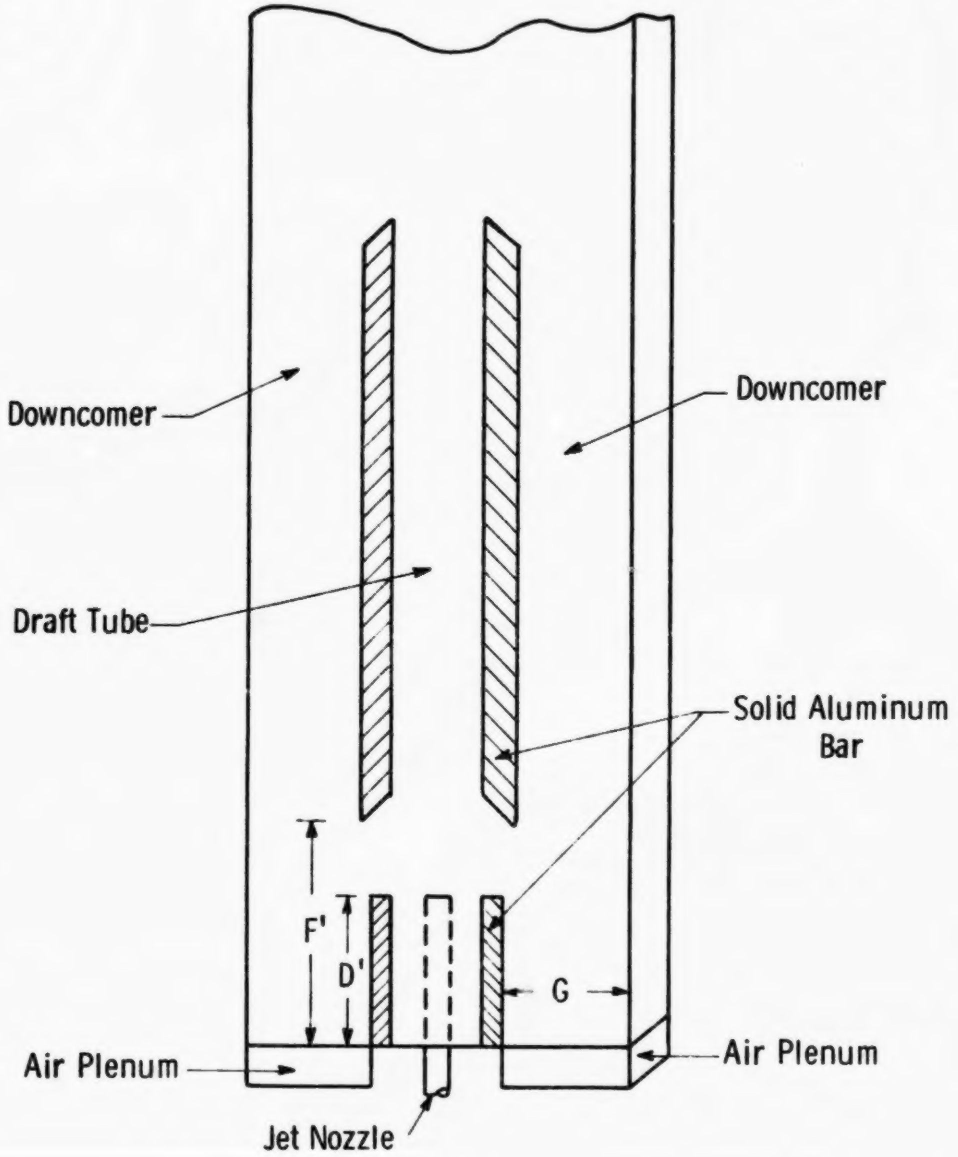
Downcomer
Downcomer
Draft Tube
Solid Aluminum Bar
F'
D'
G
Air Plenum
Air Plenum
Jet Nozzle

Figure 113—Two-Dimensional Cold Model with Modified Draft Tube Inlet

All the dimensions in the modified model are similar to the original model figure except now D' = 5 in, F' = 7 in, and G = 3.0 in.

Serpentine solid copper rods as shown in Figure 114 were also inserted in the downcomer sections to simulate the imbedded heat transfer surfaces. The serpentine rods are supported by inserting the ends of the rods into closely sized holes on the aluminum bars. Two serpentine rods were arranged in parallel 4.5 in. from the top of the draft tube with equal distances between the rods and between the rods and the walls.

Ottawa sand was used as the bed material. During operation, air is injected into the draft tube through the air jet nozzle to provide an air velocity up to ∿ 43 ft/sec in the draft tube while the downcomer sections are "minimally fluidized." This creates a solid circulation pattern upward through the draft tube and downward in the downcomers. The minimum fluidizing condition in the downcomers is assured first by turning up the air supply to the downcomers until air bubbles appear in the downcomers and then turning down the air supply until air bubbles just disappear.

The solid downward velocity in the downcomer was estimated by following a tracer particle for 16 in. with a stop watch. The tracer particles are silica gel dyed with red pigment and of similar size. For every operating condition, at least ten particles in each downcomer were traced and arithmetic average velocity was taken to be the solid particle velocity in the downcomer. The solid recirculation rate was then calculated by assuming plug flow in the downcomer.

A mathematical model has been developed to predict solid circulation rate.[12] The model is good to within ± 20% if experimental values of slugging pressure losses are used. The detailed development of the mathematical model and the experimental conditions and results are presented in Appendix P. The critical design variables for a recirculating fluidized bed boiler are identified as the draft tube inlet design, the draft tube height, and the downcomer/draft tube area ratio.

Curve 653461-A

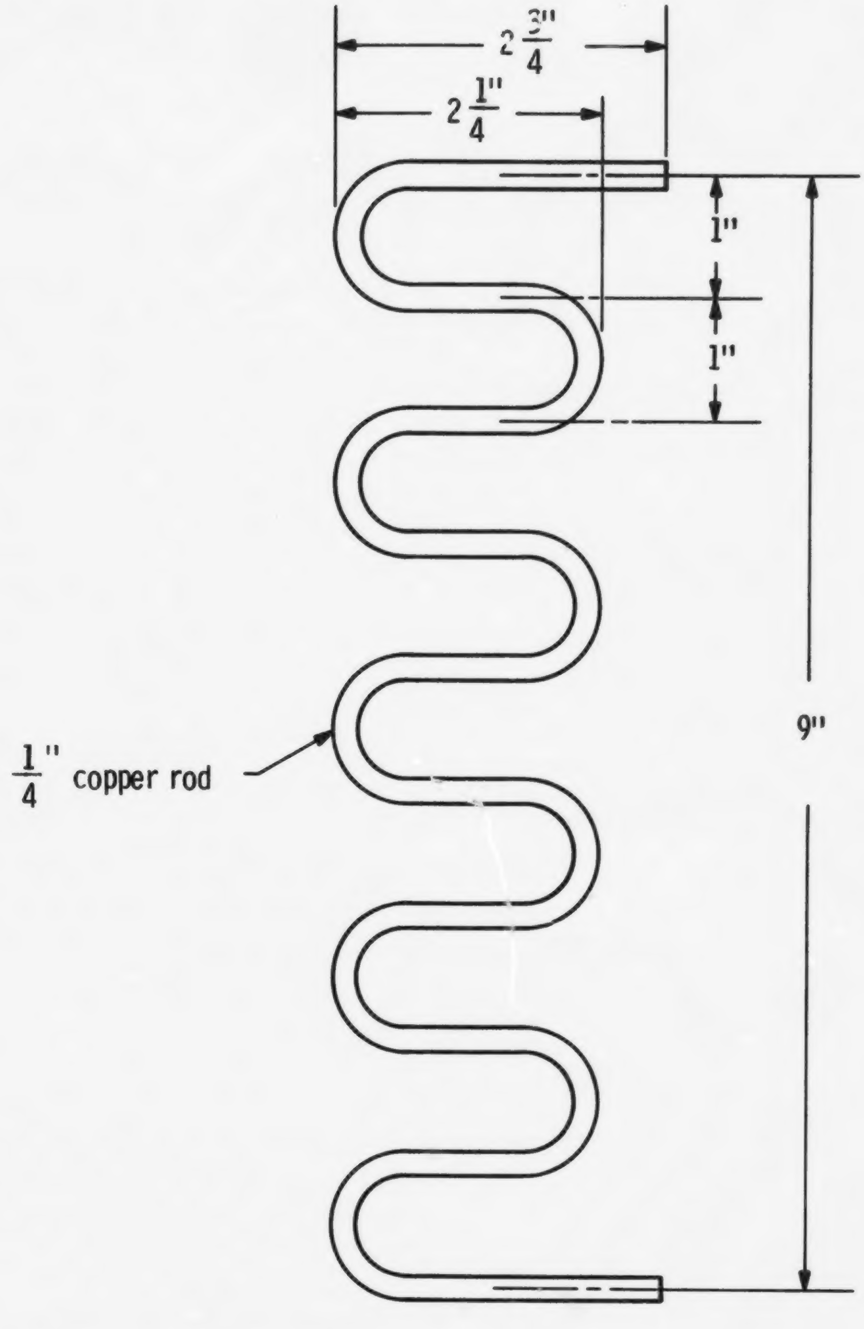
2 $\frac{3}{4}$"
2 $\frac{1}{4}$"
1"
1"
9"
$\frac{1}{4}$" copper rod

Figure 114—Dimensions of the Serpentine Copper Rods

6.2.3 Bed-Tube Heat Transfer Coefficient

No heat transfer experiment has been performed to obtain the bed-tube heat transfer coefficient for the boiler tubes immersed in the downcomer; however, literature data on heat transfer coefficient in a moving bed were utilized to project a reasonable value for the present recirculating bed boiler design. The aspects of heat transfer in a gas fluidized bed has been reviewed previously.[1] The present focus is on literature data which are relevant to heat transfer in a moving bed.

If radiative heat transfer can be ignored, the dominating factor in the heat transfer is the high heat capacity of solid particles relative to that of gas. Heat is primarily transferred from the bulk of the bed by solid convection, and convective heat transport through the gas is relatively unimportant. Because of the high surface area exposed by the particles, the solids effectively act as a local heat source or heat sink. Consequently, the thermal gradient between the immersed transfer surface and the bed is restricted to the zone immediately adjacent to the surface. Thus highest overall heat transfer coefficient will occur under conditions which limit particle residence times at the surface. It is difficult to rely on the ordinary bubbling processes in a fluidized bed for controlling the heat transfer in the bed. A forced particle circulation such as in a recirculating bed design will be able to control the particle flow past the heat transfer surfaces in a regulated manner.

Botterill et al,[13-16] examined the relation of heat transfer efficiency with particle residence time on the heat transfer surfaces both experimentally and theoretically and found that high rates of heat transfer could be obtained if particle residence time at the transfer surface was sufficiently limited. The experiments were carried out in a stirred packed bed,[13,14] and a flowing bed with a small exposed wall heat transfer surface. The bed-tube heat transfer coefficients obtained ranged from 70 to 280 Btu/ft^2-hr-°F depending on particle sizes, particle properties, and particle residence times at the heat transfer surface.

Actual heat transfer coefficients for various densities of fluidized coal were also measured. The results ranged from 80 Btu/ft^2-hr-°F for a residence time of ∿ 50 milliseconds to 190 Btu/ft^2-hr-°F for a residence time of ∿ 7 milliseconds.

Experiments were also carried out by mounting arrays of tubes on suitable jigs in the straight section of the continuous channel containing a flowing fluidized bed.[16,17] Both horizontal and vertical tube bundles were used in the experiments. Local heat transfer coefficients between elements of the surface and the bed in cross flow increased with an increasing solids flow rate, as expected. The local heat transfer coefficients are between 100 and 130 Btu/ft^2-hr-°F. However, when the solids flow rate was increased, a wake developed from the downstream -- facing surfaces, and there was a marked tendency for the bed to defluidize locally at the upstream-facing surface. Consequently, the overall bed-tube heat transfer coefficient was not very sensitive to solids flow rate and was of the order of 55 Btu/ft^2-hr-°F. There was no attempt to align the axes of the tube bundles in the flow direction of the flowing fluidized bed.

Another similar heat transfer phenomenon is that of heat transfer between submerged heat transfer surfaces and the bed in a spouted bed. The heat transfer coefficient measured in the annular section between the spout and the containing wall were found to change from ∿ 30 Btu/ft^2-hr-°F close to the wall to ∿ 50 Btu/ft^2-hr-°F close to the spout.[18] The only difference from the heat transfer in the downcomer section of a recirculating bed is that the downcomer section is minimally fluidized in the former design while there is no gas input to the annular section in the latter case. Due to this added agitation, the heat transfer coefficient in a recirculating bed is expected to be higher than that in a spouted bed.

Zenz and Othmer[19] reviewed the heat transfer in a moving bed. The typical heat transfer coefficient obtained for coal and limestone in a horizontally flowing bed of solids ranges from 40 to 100 Btu/ft^2-hr-°F.

For a moving bed with little relative movement between particles, the thermal conductivity of the solid particles becomes the controlling factor. Equations are available for these calculations.

Based on this literature review, a commercial recirculating bed boiler is designed by assuming the overall heat transfer coefficient in the downcomer section as 50 and 75 Btu/ft^2-hr-°F. The boiler costs are then compared with that of the basic design.

6.2.4 Conceptual Recirculating Bed Boiler Design

A conceptual recirculating bed boiler design is prepared on the basis of the boiler plant design developed by Westinghouse under contract to EPA.[11] The operating conditions and design parameters for a 318 MW boiler and the power cycle have been summarized in Table 5 of Section 2.1. The conceptual recirculating bed boiler design is evaluated in two cases. Case I assumes the overall heat transfer coefficient in the downcomer section to be 75 Btu/ft^2-hr-°F; and Case II assumes 50 Btu/ft^2-hr-°F. The coal is assumed to combust mostly in the draft tube and partially in the fluid bed above the draft tube. The heat transfer surfaces in the fluid bed keep the bed temperature essentially at 1750°F. The heat transfer surfaces in the downcomer section reduce the solids temperature from 1750°F at the top of the downcomer down to $\sim$ 1000°F at the bottom of the downcomer. The solids are then picked up in the draft tube to increase their temperature to 1750°F.

The design selected for a 318 MW recirculating bed boiler consists of six individual modules of 12-ft inside diameter with one module for preheater, two modules for evaporator, two modules for superheater, and one module for reheater. This design was selected to permit standard shop-fabrication of each vessel. A four-module design could also be developed, which would increase the diameter of two of the reactors. The conceptual design is depicted in Figure 115. Coal is fed concentrically with 90 to 95% stoichiometric air into the draft tube. Most of the coal is expected to be consumed in the draft tube. The

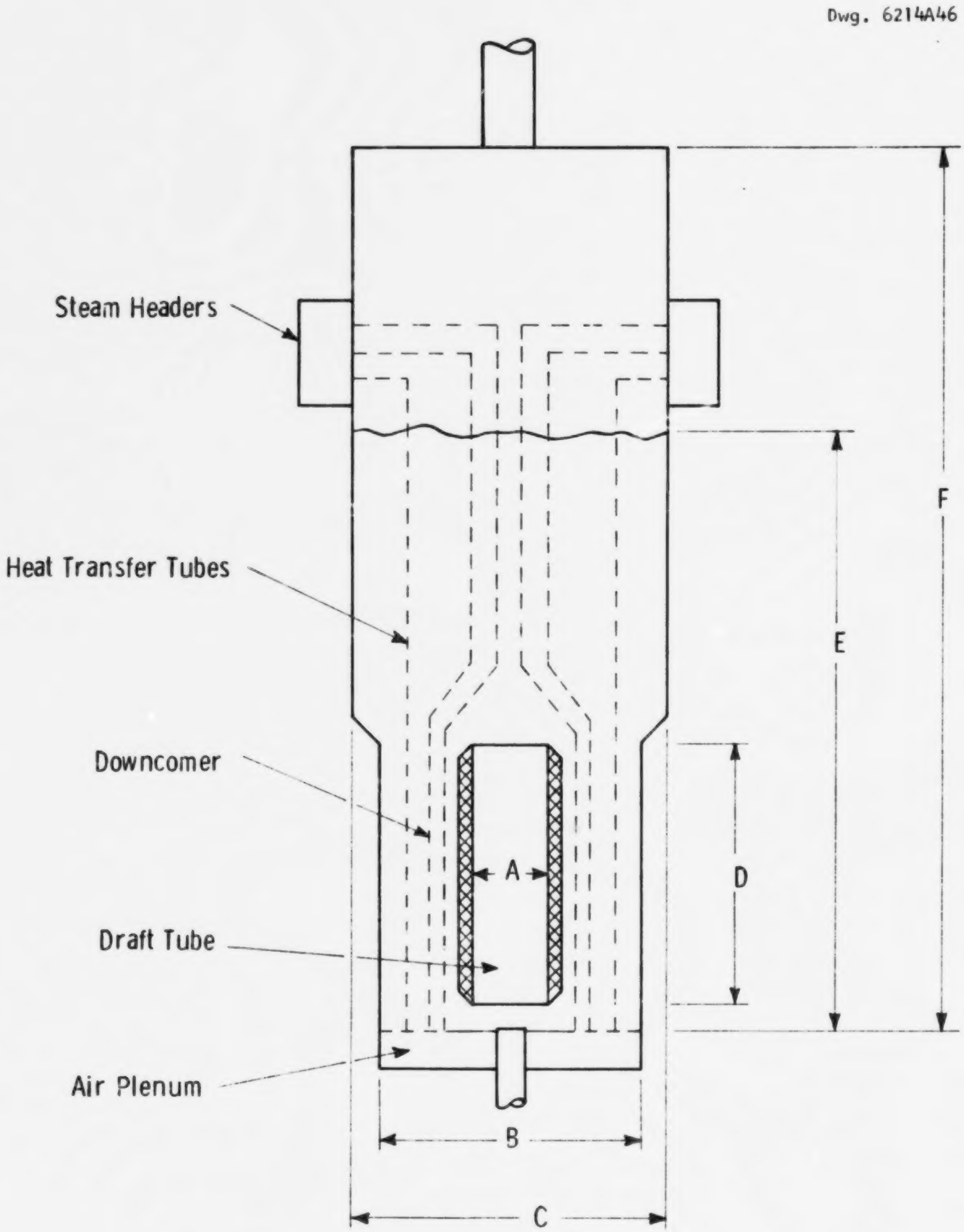
Dwg. 6214A46
Steam Headers
Heat Transfer Tubes
Downcomer
Draft Tube
Air Plenum
A
B
C
D
E
F

Figure 115 – Schematic Design of the Advanced Recirculating Bed Combustor

remaining coal is carried into the expanded fluid bed above the draft tube where additional air is available from the downcomer section. Thus the draft tube is operated under a slightly reducing condition while the fluid bed above the draft tube is operated under an oxidizing condition with 10% excess air. High excess air can be utilized in this design.[20] However, high excess air (e.g. 100%) has not been considered in the present evaluation. The effect on vessel design and performance (e.g. turndown, NO_x) would have to be considered. The bed cross-section expands at the outlet of the draft tube to reduce the superficial gas velocity to 3 to 4 ft/sec to prevent slugging and to reduce carry-over of solid particulates. The heat transfer surfaces are vertical tubes of 2 in. in diameter and extend through the downcomer section and the fluid bed above it. The horizontal bends into the steam headers just above the fluid bed will help prevent slugging and reduce elutriation. The design parameters and operating conditions are summarized in Table 54. The resulting physical dimensions of the modular combustors are presented in Table 55. The largest vessel diameter was restricted to 12-ft inside diameter so that the vessels can be shop fabricated and shipped by rail.

6.2.5 Boiler Operation and Performance

The coal-burning recirculating fluidized bed boiler would be operated somewhat differently from the basic pressurized fluidized bed boiler because of the differences in boiler design and modular arrangement. In the recirculating fluidized bed boiler design, each individual module performs only one function, i.e., it is either for preheating, evaporating, superheating, or reheating. Thus, this concept requires a minimum of four modules to complete the total boiler function from preheating to superheating and reheating. In the pressurized fluidized bed boiler design, however, each module contains one preheater bed, two superheater beds, one reheater bed, and walls for evaporating. Thus, each module is capable of generating 280 MW electrical power. Due to this difference in modular arrangement, the philosophy for cold start-up, hot restart, turndown, and load control would be different.

TABLE 54

DESIGN PARAMETERS AND OPERATING CONDITIONS FOR THE 318 MW RECIRCULATING FLUIDIZED BED BOILER

	Preevaporator	Evaporator	Superheater	Reheater
Number of Modules	1	2	2	1
Fluid Bed Temperature, °F	1750	1750	1750	1750
Temperature Change in Downcomer °F	1750→1000	1750→1000	1750→1350	1750→1350
Heat Duty, Btu/hr	252.5×10^6	616.0×10^6	636.0×10^6	323.0×10^6
Solid Circulation rate, lb/hr	1.35×10^6	1.65×10^6	3.18×10^6	3.23×10^6
Heat Transfer Surface, ft^2	6140	7140 X 2 = 14280	9320 X 2 = 18640	9800
Gas Velocity in the Draft Tube, ft/sec.	50	60	60	60
Air Feed to the Draft Tube, % stoichiometric	92	94.5	94.4	94
Superficial velocity in the fluid bed, ft/sec.	3.3	3.7	3.8	3.9
Excess air, %	10	10	10	10
Heat Transfer Coefficient, Btu/ft^2-hr-°F	75 in downcomer and 50 in bed above			
Draft Tube and Open Downcomer Area Ratio	1:10	1:10	1:10	1:10
Amount of Downcomer Area Occupied by the Heat Transfer Tubes, %	20	25	25	25

TABLE 55

PHYSICAL DIMENSIONS OF THE COMBUSTOR MODULES

Dimensions in Ft. (refer to Fig. 115)	Preevaporator	Evaporator	Superheater	Reheater
A (diameter)	2.7 i. d.	2.8 i. d.	2.8 i. d.	2.8 i. d.
B (diameter)	9.9 i. d.	10.0 i. d.	10.2 i. d.	10.3 i. d.
C (diameter)	12.0 i. d.	12.0 i. d.	12.0 i. d.	12.0 i. d.
D (height)	10.0	10.0	10.0	10.0
E (height)	18.2	15.0	19.6	19.6
F (height)	30.0	30.0	30.0	30.0

The results from the cold model studies in the two-dimensional model indicate that the solid circulation rate in a recirculating bed can be easily controlled by turning down the air input into the downcomer section. The circulation can be effectively stopped by completely shutting off the air input into the downcomer. Hence, theoretically, the recirculating fluidized bed boiler will be able to follow the load change continuously from no load to full load. Subdivision into modules for turndown consideration is not necessary. The design of the recirculating bed boiler is thus based on other considerations, such as ease in assemblage and erection, control and turndown, and economics. To be conservative, the 318 MW recirculating fluidized bed boiler design consists of six modules: one for preheating, two for evaporating, two for superheating, and one for reheating. The downcomer section of preheater and reheater modules will be separated into two sections by water walls to facilitate turndown. Air plenums for evaporator and superheater will be sectioned into several sections so that the circulation rate of each section can be individually adjusted.

The bed temperature in the fluid bed above the draft tube is the primary variable for control: too high a temperature will cause ash to fuse and dolomite to lose reactivity; too low a temperature will decrease substantially the reactivity of dolomite and possibly decrease the combustion efficiency of coal in the bed. If carbon is elutriated from the bed, it can be recycled and fed at a point just above the downcomer section so that the elutriated carbon particles can be trapped in the descending bed in the downcomer to increase residence time of the particles. There is no fresh coal feed in the downcomer section to compete for the air, and there is an additional chance for the unburned carbon particles to be picked up in the draft tube and carried into the fluid bed above the draft tube. Thus, the carbon burn-up cell in the pressurized fluidized bed design* probably will not be needed in the recirculating fluidized bed design, which would simplify the control scheme and boiler design.

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*A carbon burn-up unit is not considered necessary in either design with high excess air.

Air input into the downcomer section will be used to control the solid recirculation rate and thus the heat transfer rate to the heat transfer surfaces. Because of this great flexibility, cold start-up, hot restart, and load control are much simpler. To start up, auxiliary fuel can be fed into the fluid bed through the draft tube. When the bed temperature reaches the coal ignition point ($\sim$ 1000°F), coal feed can be started. The solid circulation can be started when the bed temperature reaches the design temperature 1750°F. For hot restart, the recirculating bed design would be even more convenient, because the bed is not completely shut off during turndown as in the case of the pressurized fluidized bed design. The bed is essentially maintained at the operating temperature at least for some sections. A hot restart will simply mean an increase in the coal feed rate and the solid circulation rate, while in the pressurized fluidized bed design, the bed has to be brought up to the coal ignition temperature by auxiliary firing before coal feeding can be initiated. This means an additional time delay in following the load change. This potential capability of shortening the load response time is an important advantage for the recirculating fluidized bed design.

The performance characteristics of the recirculating fluidized bed are not expected to be too different from those of the basic pressurized fluidized bed design. The overall boiler efficiency is expected to be comparable, with slightly higher pressure drop through the recirculating fluidized bed due to deeper bed design; however, the air pollution control capability in the recirculating bed is potentially greater. The two-stage combustion of the coal in the recirculating bed -- reducing conditions in the draft tube section and oxidizing conditions in the downcomer section and the main bed -- promises to reduce the formation of NO_x. If fluctuation of the sorbent between oxidizing and reducing conditions promotes penetration of the sulfur into the sorbent in the form of sulfate, sulfite, or sulfide, then increased activity and higher utilization of the sorbent can be achieved.

6.2.6 Economics

In addition to the air pollution control capability, the recirculating fluidized bed boiler possesses the following potential advantages:

- In decreasing coal feed points. Coal can be fed and combusted in the draft tube through a single coal feed point. The dilute pneumatic transport of coal particles in the draft tube increases gas/solid contact. The air by-passing through bubbles which adversely affect the gas/solid contact in a fluidized bed does not occur in the draft tube.
- In preventing agglomeration. High solid circulation rate in the recirculating bed prevents caking and agglomeration of coal in the vicinity of coal feeding points.
- In capability of designing for deeper bed. The increased circulation and mixing in a recirculating bed eliminates the problems of particle segregation and temperature gradient in the main bed. This permits design for a deeper bed.
- In control and load response. Continuous turndown is possible in a recirculating bed. For cold start-up or hot restart, a recirculating bed is easier to handle. The response time is shorter in following load change in a recirculating bed as discussed in the preceding section.
- Possibly in increasing the heat transfer coefficient in the downcomer section over that in a fluidized bed. Because of fast solid circulation through the downcomer section, there is a possibility of increasing the heat transfer coefficient in the downcomer section to higher than that in a fluidized bed (50 Btu/ft^2-hr-°F). This possibility has to rely on experimental verification.

All the advantages mentioned here can be directly or indirectly transferred into economic savings in designing a recirculating fluidized bed boiler.

The current design for the 318 MW recirculating fluidized bed boiler is a six-module plant with one module for preheating, two for evaporating, two for superheating, and one for reheating. Each module consists of only one deep recirculating bed as compared to four stacked beds in the pressurized fluidized bed boiler design. The pressurized vessels are designed for operation at maximum operating conditions of 150 psig and 650°F with 6-in. refractory insulation (4-in. insulation plus 2-in. hard face). Draft tubes are constructed from either 316 stainless steel or Incoloy 800 and are to be water- or steam-cooled. Vertical tubes of conventional boiler tube materials are used for heat transfer surfaces. The total cross-section of the reactor vessel is used for the recirculating bed area as compared to < 40% utilization of vessel cross-section area in the pressurized fluid bed design. The cost of the six-module boiler is evaluated for the heat transfer coefficient in downcomer the section of 75 and 50 Btu/ft^2-hr-°F. The heat transfer coefficient in the fluid bed above the draft tube is assumed to be 50 Btu/ft^2-hr-°F. The heat transfer surface required in the recirculating bed design and the boiler cost are compared with those of the basic design in Table 56. Although the heat transfer surface requirements for the recirculating bed design are 6% to 38% higher than those for the basic pressurized fluidized bed design, the overall boiler costs may be lower by 5% to 13%. The increase in heat transfer surface requirements is due to a decrease in temperature difference between the descending bed in the downcomer section and the heat transfer surface; however, the decrease in module height from four modules of 110 ft to six modules of 30 ft substantially reduces the boiler costs. This reduction in module height is primarily due to elimination of a stacked bed design. When fluidized beds are stacked one on top of the other as in the basic design, freeboard and air plenum have to be provided for each bed, thus increasing the module height.

TABLE 56

COMPARISON OF HEAT TRANSFER SURFACE REQUIREMENTS AND BOILER COSTS

	Basic Design	Recirculating Bed Design	
Heat Transfer Coefficient in Downcomer, Btu/ft^2-hr-°F	---	75	50
Heat Transfer Surfaces, ft^2			
Preevaporator	9,156	6,140	7,830
Evaporation	12,684	14,280	19,060
Superheater	15,944	18,640	24,240
Reheater	8,280	9,800	12,300
TOTAL	46,064	48,860	63,430
Boiler Cost	$3,856,000 [a]	$3,370,000 - $3,488,000 [b]	$3,537,000 - $3,670,000

[a] Cost is for a 318MW 4-module shop-fabricated unit.

[b] Cost is for a 318MW 6-module shop-fabricated unit. The price range indicates the cost differential in construction between 316 stainless steel draft tube - tube cooled wall versus Incoloy draft tube - refractory insulated wall.

The cost of the particulate removal system is expected to be comparable in both designs because the airflow rate is similar and the dust loading is expected to be similar, as well. Lower velocities in the freeboard area in the recirculating bed design should result in slightly lower dust loading if slugging does not set in.

The structural cost for six modules, each 30 feet high, should be lower than that for four modules, each 110 feet high. The manifolding, however, would be more expensive for the 6-module design. The instrumentation and control are projected to be easier and less costly for the recirculating bed design. Only 6 beds are involved, as compared to 16 individual beds in the pressurized fluidized bed design. It requires turning on and off a complete fluidized bed or a complete module in the pressurized fluidized bed for turndown purposes, while in the recirculating bed design it only requires adjustment in downcomer airflow rate. The coal feeding points are reduced from 64 to 6, which also substantially reduces the boiler costs. Thus, it is probably conservative to assume that the cost of auxiliary systems are comparable for both the recirculating bed and the pressurized fluidized bed designs.

6.2.7 Conclusions

The overall boiler costs of a recirculating bed design are estimated to be lower by 5% to 13% than those of the pressurized fluidized bed basic design, depending on the construction of the draft tube and on the heat transfer coefficient in the downcomer section. The costs are arrived at by assuming that the costs of the auxiliary systems are comparable. This is a conservative assumption, in view of the simplicity in operating a recirculating bed boiler. In addition to the cost advantage, the recirculating bed design has the following potential:

- In preventing agglomeration. High solid circulation rate in the recirculating bed prevents caking and agglomeration of coal in the vicinity of coal feeding points so that low-grade coal can be used.

- In capability of deep bed operation. The increased circulation and mixing eliminates the problems of particle segregation and temperature gradient in the main bed. This permits design for a deeper bed.
- In control and load response. Continuous turndown without shutting off the bed is possible. This promises easier start-up and faster response in following load changes.
- In SO_2 removal and NO_x reduction. The two-stage combustion -- operating the riser portion of the bed under the reducing conditions and the main bed and downcomer section under oxidizing conditions -- reduces the formation of NO_x and promotes activation of the sorbent.

The recirculating boiler concept offers many potential advantages over the base pressurized fluid bed boiler design and should be developed further. However, the recirculating bed concept has not been demonstrated on a sufficient scale to permit commercialization prior to the base design. Further information on heat transfer coefficients and testing of the heat transfer surface design, air distribution design, solid circulation control, and fluidization characteristics are required. Thus, a recirculating fluid bed boiler offers economical and environmental advantages as a second generation pressurized fluid bed boiler.

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